PB81-213274

Feasibility Study of Open Tank Oxygen-Activated Sludge Wastewater Treatment

FMC Corp.
Englewood, CO

Prepared for

Municipal Environmental Research Lab.
Cincinnati, OH

Jun 81

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1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
EPA-600/2-81-095	ORD Report	PB81 213274

4. TITLE AND SUBTITLE	5. REPORT DATE
FEASIBILITY STUDY OF OPEN TANK OXYGEN-ACTIVATED SLUDGE WASTEWATER TREATMENT	June 1981
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)	8. PERFORMING ORGANIZATION REPORT NO.
Kerwin R. Rakness*	
9. PERFORMING ORGANIZATION NAME AND ADDRESS	10. PROGRAM ELEMENT NO.
FMC Corporation 3999 South Mariposa Street Englewood, Colorado 80110	AZB1B, D.U.B113,Task No.D-1/15
	11. CONTRACT/GRANT NO.
	Grant No. S801790
12. SPONSORING AGENCY NAME AND ADDRESS	13. TYPE OF REPORT AND PERIOD COVERED
Municipal Environmental Research Laboratory -Cin.,OH Office of Research and Development U.S. Environmental Protection Agency Cincinnati, Ohio 45268	Final, May 1973-October 1974
	14. SPONSORING AGENCY CODE
	EPA/600/14

15. SUPPLEMENTARY NOTES

Project Officer: Richard C. Brenner (513)684-7657
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16. ABSTRACT

The pilot plant for this study consisted of one oxygenation basin and two clarifiers. The system treated primary clarifier effluent from the Englewood, Colorado, treatment facility. The influent flow rate was adjusted to attain average aeration reactor detention times ranging from 0.94 to 3.3 hr. The pilot plant operation was conducted in two phases. Treatment performance during both phases was excellent. Final effluent BOD_5 concentrations and secondary BOD_5 removals averaged less than 20 mg/l and greater than 90 percent, respectively. No degradation in process removal efficiency occurred even at organic loadings as high as 1.23 kg BOD_5 applied/day/kg MLVSS (1.23 lb BOD_5 applied/day/lb MLVSS) and volumetric loadings as high as 4.07 kg BOD_5 applied/day/m^3 (254 lb BOD_5 applied/day/1000 ft^3) of reactor capacity.

Analysis of pilot plant operations indicated that somewhat less sludge was produced with the oxygen system when compared to literature-cited, typical air sludge production. This decreased sludge production occurred and became more pronounced at higher organic loadings. Comparisons were made between literature-cited, typical air-sludge and oxygen-sludge settling characteristics. In all cases, the oxygen sludge initial settling velocity was greater than typical air sludge at given TSS concentrations. Average oxygen utilization efficiency was 92 percent.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
*Sewage treatment, *Activated sludge process, Sludge, Clarification	*Oxygen-activated sludge, Loading rates, Sludge production, Sludge settleability	13B

18. DISTRIBUTION STATEMENT	19. SECURITY CLASS *(This Report)*	21. NO. OF PAGES
Release to Public	Unclassified	65
	20. SECURITY CLASS *(This page)*	22. PRICE
	Unclassified	

EPA Form 2220-1 (9-73)

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EPA-600/2-81- 095
June 1981

PB81-213274

FEASIBILITY STUDY OF OPEN TANK OXYGEN-ACTIVATED SLUDGE WASTEWATER TREATMENT

by

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Grant No. S801790

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

This report describes a feasibility study of an open tank oxygen-activated sludge treatment system. The efficiency of a fine bubble diffuser is evaluated.

Francis T. Mayo
Director
Municipal Environmental Research
Laboratory

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ABSTRACT

Early investigation of high purity oxygen activated sludge treatment systems concluded that oxygen activated sludge treatment was economically attractive when a high percentage (90 percent) of the oxygen supplied was utilized. Two types of oxygen activated sludge systems are available to date: a "closed tank" system, and an "open tank" system. The open tank system discussed in this report provided the high percentage (90 percent) oxygen utilization to the mixed liquor through an ultra fine bubble diffuser. The pilot plant used the fixed active diffuser (FAD), which will be described in this report.

The pilot plant for this study consisted of one oxygenation basin and two clarifiers. The system treated primary clarifier effluent from the Englewood, Colorado, treatment facility. The influent flow rate was adjusted to attain average aeration reactor detention times ranging from 0.94 to 3.3 hr. The pilot plant operation was conducted in two phases. Treatment performance during both phases was excellent. Final effluent BOD_5 concentrations and secondary BOD_5 removals averaged less than 20 mg/l and greater than 90 percent, respectively. No degradation in process removal efficiency occurred even at organic loadings as high as 1.23 kg BOD_5 applied/day/kg MLVSS (1.23 lb BOD_5 applied/day/lb MLVSS) and volumetric loadings as high as 4.07 kg BOD_5 applied/day/m^3 (254 lb BOD_5 applied/day/1000 ft^3) of reactor capacity.

Analysis of pilot plant operations indicated that somewhat less sludge was produced with the oxygen system when compared to literature-cited, typical air sludge production. This decreased sludge production occurred and became more pronounced at higher organic loadings. Comparisons were made between literature-cited, typical air-sludge and oxygen-sludge settling characteristics. In all cases, the oxygen-sludge initial settling velocity was greater than typical air sludge at given TSS concentrations.

The top of the oxygenation basin was covered for 4 to 5 days and a thorough evaluation of the quantity of oxygen in the "off-gas" was made on two occasions during the program. Results of these tests showed greater than 90 percent of the oxygen supplied to the system was utilized. Average oxygen utilization efficiency was 92 percent.

This report was submitted in fulfillment of Grant No. S801790 by the FMC Corporation under the partial sponsorship of the U.S. Environmental Protection Agency and covers the period May 1973 through October 1974.

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ABBREVIATIONS

Accum	Accumulated
Appl	Applied
BOD_5	Five-day biochemical oxygen demand
^oC	Degrees Celsius (centigrade)
cm	Centimeter
COD	Chemical oxygen demand
Conc	Concentration
D.O.	Dissolved oxygen
FAD	Fixed Active Diffuser (MAROX)
ft	Foot or feet
FE	Final effluent
FTU	Formazin turbidity units
^oF	Degrees Fahrenheit
F/M	Food to microorganism ratio
gpm	Gallons per minute
gal	Gallon
gm	Gram
hr	Hour
Hg	Mercury
in.	Inch or inches
JTU	Jackson turbidity units
kg	Kilogram
l	Liter
lb	Pound
m	Meter
Max	Maximum
ml	Milliliter
mg	Milligram
Min	Minimum
ML	Mixed liquor
MLSS	Mixed liquor suspended solids
MLTSS	Mixed liquor total suspended solids
MLVSS	Mixed liquor volatile suspended solids
NH_3-N	Ammonium nitrogen
NO_2-N	Nitrite nitrogen
NO_3-N	Nitrate nitrogen
No.	Number
psia	Pounds per square inch absolute
psig	Pounds per square inch gauge
RAS	Return activated sludge
Rem	Removed
R/Q	Return flow/influent flow
scfm	Standard cubic feet per minute
Std	Standard
SI	Secondary influent
Supp	Supplied
SVI	Sludge volume index
TKN	Total Kjeldahl nitrogen
TSS	Total suspended solids
VSS	Volatile suspended solids
WAS	Waste activated sludge

ACKNOWLEDGMENTS

This project was funded by the Environmental Protection Agency, Grant No. S801790. Mr. Richard C. Brenner, Municipal Environmental Research Laboratory (MERL), Cincinnati, Ohio, was the Project Officer. Mr. Richard L. Fowle, also of MERL, reviewed this report.

The project was administered by MAROX Diffusion Systems staff. Mr. Richard B. Weber was the Grant Director, and Mr. Duane M. Parker was Project Director. Principal field administrators for the MAROX operations were Mr. Michael Joyce and Mr. Kerwin Rakness.

The cooperation and assistance of the staff of the Englewood, Colorado, Treatment Facility helped considerably in the monitoring and laboratory support requirements of this project. In this respect, Mr. Charles Carroll, Director of Public Utilities, and Mr. Jack Reid, Superintendent of the Englewood Treatment Facility, are specifically recognized for their able assistance.

SECTION 1

INTRODUCTION

GENERAL BACKGROUND

The investigation of high purity oxygen for sewage treatment was originally conducted by Okun in 1948.[1] No significant further development was conducted until the late 1960's when a "closed tank" oxygenation process was developed.[2] The closed tank process demonstrated that oxygen activated sludge treatment was feasible and economically attractive when compared to a parallel operated air-activated sludge system. Two of the reasons for the attractiveness of the closed tank process were: (1) ability to treat the wastewater with smaller aeration reactors than required for air systems, and (2) ability to achieve high (90 percent) oxygen utilization efficiencies.

There are many reasons why an "open tank" oxygenation system is attractive. These reasons are: (1) extensive safety precautions are required with the closed tank system; (2) extra expense is involved in covering and sealing an aeration reactor as well as a requirement for 0.91 to 1.5 m (3 to 5 ft) of freeboard; (3) high purity oxygen activated sludge treatment has been shown to be feasible and economically attractive at high oxygen utilization efficiencies; and (4) existing aeration basins may be easily converted to oxygen systems in an open tank concept. To date there is one known open tank high purity oxygen activated sludge system on the market. This system was originally developed by FMC Corporation under the trade name of MAROX. Rights to the MAROX technology have since been acquired by Zimpro Inc. The basis for performance of the MAROX system is an ultra fine bubble diffuser that achieves oxygen utilization efficiencies comparable to those obtained with closed tank reactor systems.

MAROX Diffuser Technology

In order to achieve high oxygen utilization rates in an open basin, 90 percent of all oxygen supplied must be transferred to the mixed liquor. A paper written by Richard B. Weber[3] (see Appendix) describes in detail the factors involved in achieving high oxygen utilization rates with the MAROX process.

The MAROX system uses a unique bubble shear method for dissolution of pure oxygen to the mixed liquor. The Fixed Active Diffuser (FAD) hardware used in this project had two gas bars adjacent to each slot that emitted oxygen through very small capillaries where a high velocity mixed liquor stream sheared the bubbles on formation into 50- to 100-micron-diameter gas bubbles (Figure 1). The mixed liquor stream flows perpendicular to the gas emitting surface and is pumped by an externally mounted pump for each pass

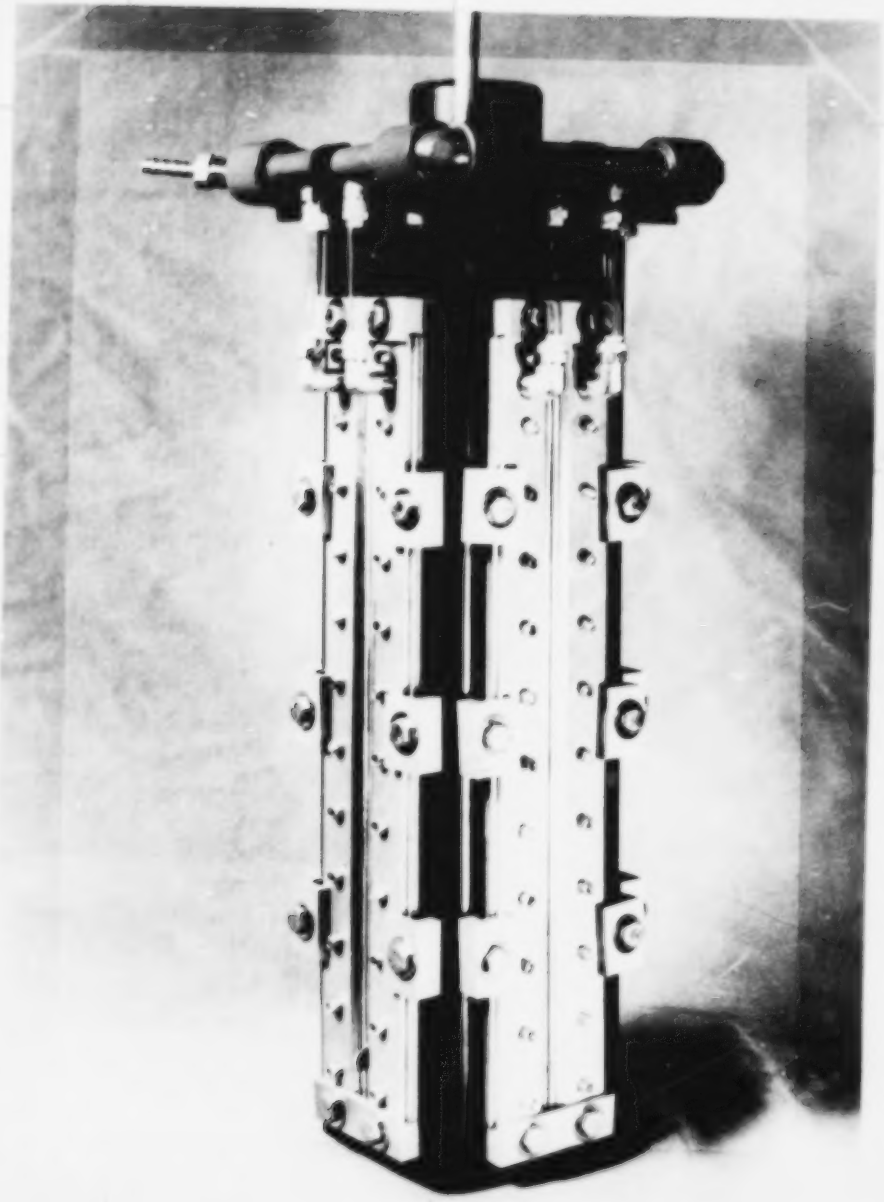

Figure 1. Fixed active diffuser.

recycling the mixed liquor in that particular pass. The mixed liquor flow through each of two vertical slots of each diffuser provided both shear and mixing requirements for the pass. The slot width of the diffuser required prescreening of the secondary influent to eliminate solids larger than 1.02 mm (0.040 in) from entering the mixed liquor basin.

Treatment Plant Description

The Englewood, Colorado treatment facility, where the MAROX pilot plant operated, received approximately 32,000 m^3/day (8.5 mgd) of sewage from residential, commercial, and industrial sources. The primary source of the plant flow is residential; a small fraction is from commercial and industrial sources. Three industrial sources that impact the characteristics of the Englewood raw sewage are a dairy products processing plant, a metal plating firm, and a septic tank cleaning firm. No attempt will be made to quantify the impact of these industrial sources on the Englewood raw sewage characteristics, but the relative impact can be noted in the occasional higher than normal BOD and total suspended solids (TSS) concentration of the raw sewage. The metal plating firm on one occasion discharged toxic waste to the treatment plant during this program, but no significant impact was noted in the pilot plant operation. The raw sewage flow entering the Englewood treatment facility receives coarse screening and grit separation. After grit separation the flow splits into two streams (see Figure 2 for schematic diagram). The south stream provides the influent source for the pilot plant. Sewage flow to the south plant is measured using a Parshall flume with pumping required to the primary clarifier. Primary effluent is then treated in two high-rate trickling filters after which a contact stabilization activated sludge system provides a further upgrading of the effluent. The effluent streams of the north and south system join at the chlorine contact basin for chlorination prior to discharge into the South Platte River.

Sludge handling at the Englewood treatment facility consists of anaerobic digestion followed with sludge drying beds. Waste activated sludge (WAS) from the south plant contact stabilization unit is recycled to the south plant primary clarifier. Supernatant from the south plant and north plant second stage anaerobic digesters is also recycled to the south plant primary clarifier. These two in-plant recycle flow streams had a significant impact on the primary clarifier effluent and the pilot plant influent.

Primary clarifier effluent from the south plant was either gravity drained or pumped to the MAROX pilot plant during the course of this project. During those periods of operation when the pilot plant flow rate was greater than 114 l/min (30 gpm), the pilot plant sewage influent was pumped to the pilot plant. During all other periods of operation the pilot plant sewage influent was gravity drained to the pilot plant.

Program Plan

The following objectives were established to verify the performance of the pure oxygen open tank activated sludge system: (1) determine the final effluent quality, process removal efficiency, and sludge production characteristics at various secondary influent flow conditions; (2) determine oxygen

KEY:

CB - CONTACT BASIN
CS - COURSE SCREENING
DS - DIGESTED SLUDGE
DSP - DIGESTER SUPERNATANT
FC - FINAL CLARIFIER
FM - FLOW MEASUREMENT
GR - GRIT REMOVAL
PC - PRIMARY CLARIFIER
PD - PRIMARY DIGESTER
PS - PRIMARY SLUDGE
RAS - RETURN ACTIVATED SLUDGE
RB - REAERATION BASIN
RSWW - RAW SEWAGE WET WELL
RWW - RECIRCULATION WET WELL
SD - SECONDARY DIGESTER
SS - SECONDARY SLUDGE
TF - TRICKLING FILTER
WAS - WASTE ACTIVATED SLUDGE

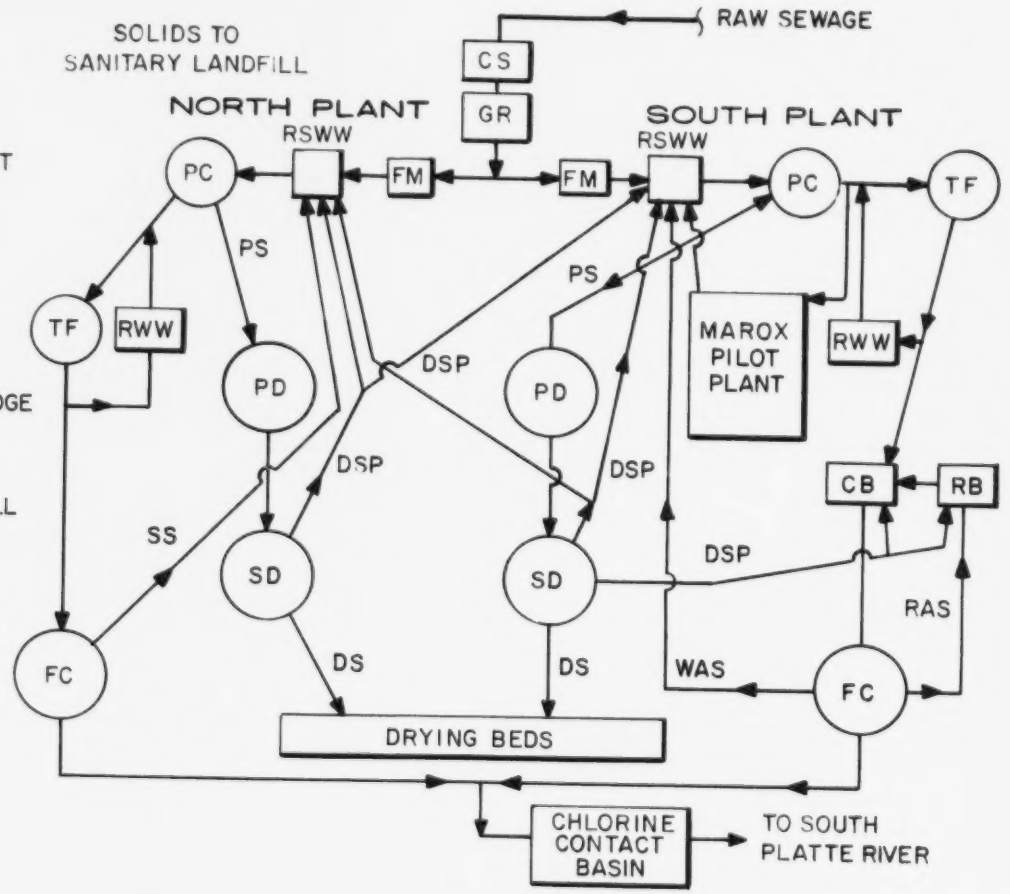

Figure 2. Schematic diagram, Englewood wastewater treatment facility.

usage efficiencies using an open tank system; (3) determine any effect of the shearing action for production of minute oxygen bubbles on the settling characteristics of the mixed liquor; (4) determine the oxygen feed control response using a D.O. monitoring and control system for oxygen supply; (5) identify mixed liquor recirculation rates required for system operation; and (6) identify power requirements for system operation.

Testing was completed using a pilot plant facility capable of treating secondary influent flow rates up to 151 l/min (40 gpm). Secondary influent (SI) to the pilot plant received primary treatment at the Englewood facility. During the testing period, one oxygenation basin and either one or two clarifiers were in service. Table 1 shows the program plan with both constant flow and the maximum-minimum diurnal flow conditions. Diurnal flow variations were selected based on the historical raw sewage flow to the Englewood facility.

TABLE 1. PROGRAM PLAN OPERATING CONDITIONS

Test No	Average Daily Flow (l/min)	Flow Condition	Maximum Daily Flow (l/min)	Minimum Daily Flow (l/min)	Reactor* Detention (hr)	Clar No	Clarifier+ Overflow Rate ($m^3/day/m^2$)	Clarifier‡ Overflow Rate ($m^3/day/m^2$)
1	37.8	Constant			3.30	1	15.2	16.7
2	37.8	Diurnal	47.3	22.7	3.30	1	15.2	16.7
3	56.7	Constant			2.20	1	22.9	25.1
4	75.8	Diurnal	102.0	45.5	1.65	1	30.5	33.5
5	94.5	Constant			1.32	2	19.1	21.0
6	113.5	Diurnal	133.5	75.8	1.10	2	22.9	25.1
7	132.5	Constant			0.94	2	26.7	29.3

Note: l/min x 0.264 = gpm
$m^3/day/m^2$ x 24.6 = gpd/ft^2

* Based on average daily SI flow.
\+ Includes clarifier centerwell.
‡ Excludes clarifier centerwell.

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SECTION 2

RESULTS AND CONCLUSIONS

The MAROX oxygen pilot plant provided data which led to the following conclusions: (1) Final effluent BOD_5 concentrations averaged less than 20 mg/l even at average volumetric loadings as high as 4.07 kg BOD_5 applied/day/m^3 (254 lb BOD_5 applied/day/1000 ft^3) of oxygenation basin capacity, and maximum organic loadings of 1.23 mass BOD_5 applied/day/mass mixed liquor volatile suspended solids (MLVSS); (2) Better effluent quality was attained by operating at return flow/influent flow (R/Q) ratios between 50 percent and 60 percent as opposed to R/Q ratios between 30 percent and 40 percent; (3) No significant nitrification occurred during the test program; oxygenation basin detention times ranged from 0.94 hr to 3.3 hr; (4) Excellent oxygen feed control response was achieved using a dissolved oxygen (D.O.) monitoring and control system to vary oxygen supply rate; (5) Oxygen utilization efficiencies averaged 91.5 percent and 92.5 percent during two separate oxygenation basin "off-gas" tests; (6) Oxygen-sludge mixed liquor suspended solids (MLSS) initial settling velocities were greater than literature-cited air-sludge settling velocities at specific TSS concentrations; (7) Oxygen-sludge production values were somewhat lower than most literature-cited air-sludge production; and (8) Final effluent pH was within 0.1 unit of the secondary influent pH.

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SECTION 3

RECOMMENDATIONS

The following recommendation is made from the results of the MAROX pilot plant operation. For oxygen utilization efficiency verification in an "open" oxygenation basin, the following approach is recommended for activated sludge systems: (1) compare ratio of mass of oxygen supplied per mass of BOD_5 removed with expected ratios. If the calculated ratio is equal to or less than this ratio, then the oxygen utilization efficiency is acceptable. If the calculated ratio is greater than the expected ratio, then the oxygen utilization efficiency may not be acceptable and further testing is required; (2) an accurate oxygen utilization efficiency can be determined by using a cap or cover on all or a portion of the surface area of the oxygenation basin and conducting an off-gas analysis. Other indirect methods for obtaining mass of oxygen used are not acceptable alternatives.

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SECTION 4

MATERIALS AND METHODS

PILOT PLANT DESCRIPTION

The MAROX high purity oxygen activated sludge pilot plant was located at the Englewood treatment facility. A complete description of the Englewood treatment system was explained in the introduction to show the position of the pilot facility within the treatment plant, and to identify the Englewood in-plant recycle flows which affected the pilot plant influent.

Figure 3 shows a schematic diagram with unit capacities of the pilot plant facility. Effluent from the Englewood primary clarifier (secondary influent) entered the top of a "fine" screen designed to remove solids greater than 1.02 mm (0.04 in.) in diameter. Screened sewage then entered a 37.85-liter (10-gallon) constant head holding tank. A greater volume of SI was screened than was directed through the pilot plant with the excess screened sewage returned to the raw sewage wet well. Screened sewage was pumped to the oxygenation basin using either one or two centrifugal pumps. The SI flow rate was varied by adjusting the opening of a ball valve. The SI flow rate was determined using a 18.93-liter (5-gallon) calibrated measuring device versus time. SI was directed to the measuring device by a three-way valve. The aeration basin was built to approximate a plug flow condition by dividing it into three separate flow through compartments connected in series. The three separate compartments were labeled "A", "B", and "C" pass. Mixed liquor traveled through the aeration basin from the top to the bottom of the "A" pass, from the bottom to the top of "B" pass, and from the top to the bottom of "C" pass. Mixed liquor was directed from the bottom of "C" pass to either one or two final clarifiers.

Both final clarifiers were identical in size. Mixed liquor entered the centerwell of each clarifier 0.61 m (2 ft) below the liquid surface and with the centerwell skirt extended 1.52 m (5 ft) below the liquid surface. A horizontal baffle was located 15.2 cm (6 in.) below each centerwell skirt for mixed liquor distribution. Each clarifier had a 20^{o} sloping bottom with the return activated sludge outlet located at the apex. Return sludge was gently moved to the outlet with a conventional scraper mechanism. Final effluent was discharged from the clarifier over a peripheral weir, back to the Englewood raw sewage wet well. Return activated sludge (RAS) from each clarifier was directed to the "A" pass of the oxygenation basin using a centrifugal pump. The RAS flow was controlled and measured in the same manner as the SI flow. Batch wasting from the system was accomplished by diverting RAS to a 189-liter (50-gallon) barrel calibrated in 37.9-liter (10-gallon) increments. During periods of high SI flow and the wasting of greater sludge quantities, RAS was continuously wasted at a low-flow for selected periods. This WAS flow

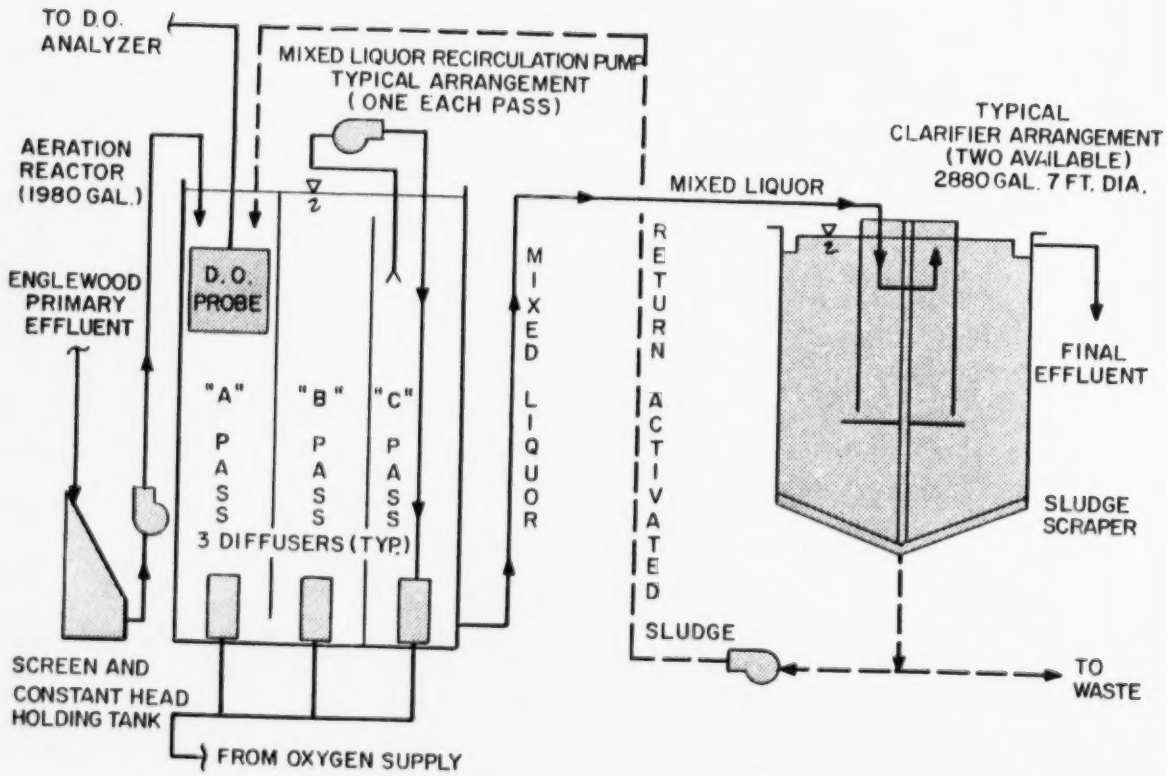

Figure 3. Elevation schematic diagram, MAROX pilot plant.

was determined using a calibrated 18.9-liter (5-gallon) measuring container versus time. In both cases, WAS was discharged to the Englewood raw sewage wet well. During the course of the project, either one or two FAD's were located in "A" pass. In all cases one FAD was used in "B" and "C" passes. Oxygen-bubble shearing and oxygenation basin mixing were accomplished using one centrifugal pump for each pass. The slotted openings of the diffusers in "A" and "B" passes were kept free from clogging by manually moving a knife blade cleaning device up and down the length of the slotted opening once per day. "C" pass was equipped with a reverse flow cleaning method to remove entrapped material once per day.

Throughout the project, a 708-m^3 (25,000-ft^3 gas equivalent) liquid oxygen storage tank with vaporizer was used. Gaseous oxygen to the diffusers was monitored through an oxygen panel where gauges, meters and valves were located to control and measure oxygen flow to each pass. Figure 4 shows the oxygen-gas flow control for the pilot plant. The system's oxygen gas pressure was adjusted with a pressure regulating valve located on the main oxygen line from which oxygen was provided for each of the three passes. Initially, oxygen supplied to each pass was varied manually by a valve located on the oxygen line to that pass. The oxygen supplied to each pass was calculated using data from the rotometer reading, oxygen pressure readings and oxygen temperature readings for each pass. Later in the project, a continuously operating D.O. probe was placed in "A" pass at 1.22 m (4 ft) below the liquid surface for automatic oxygen control with "B" and "C" pass diffusers manually adjusted. The D.O. analyzer current/pneumatic (I/P) converter, and pneumatic recorder-controller were installed in a service building about 24.4 m (80 ft) from the pilot plant. The pneumatic oxygen control valve was situated on the back of the control panel which was adjacent to the oxygenation basin.

The components of the automatic control system from the D.O. sensor to the oxygen flow control valve were as follows:

1. Weston & Stack Model A-40 dissolved oxygen probe complete with Model A-25 agitator
2. Weston & Stack Model 3000-1-A dissolved oxygen analyzer
3. Foxboro Model 69TA1 current-to-pneumatic (I/P) converter
4. Foxboro Model 121EE dissolved oxygen recorder
5. Foxboro Model 130M-N4 dissolved oxygen controller
6. Foxboro V4A series pneumatically adjusted oxygen supply control valve

The automatic oxygen flow-control functioned as a closed loop system which operated in the following manner. The D.O. probe sensed the D.O. concentration of the mixed liquor. Tha agitator assembly kept the probe's membrane free from solids accumulation and circulated "fresh" mixed liquor past the membrane surface. The D.O. analyzer converted the signal received from the probe to an electrical current and instantaneously displayed the D.O. concentration level.

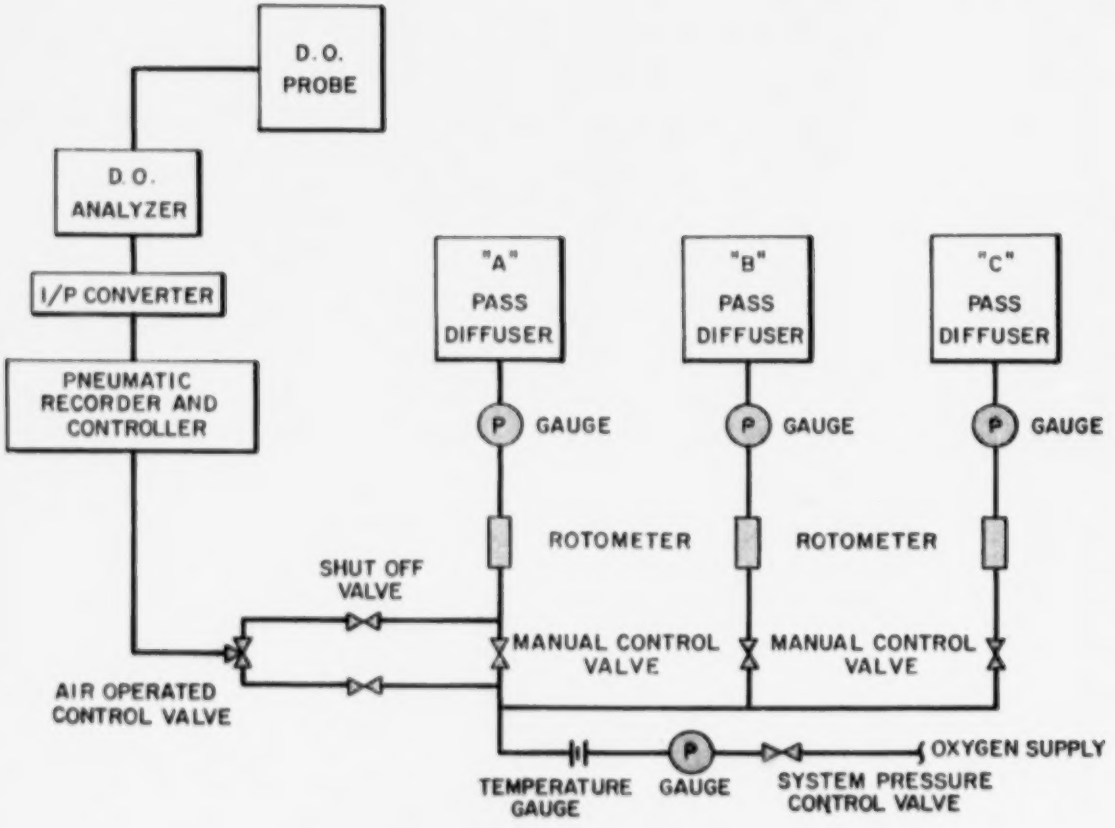

Figure 4. MAROX pilot plant oxygen flow schematic.

The analyzer also sent the electrical signal to the I/P converter which changed it into a corresponding pneumatic signal. The D.O. controller compared the pneumatic signal received from the I/P converter with a "set-point" signal that was manually adjusted to the D.O. concentration level desired. The controller then sent an appropriate pneumatic signal to the oxygen supply control valve to increase or decrease the oxygen flow to correspond with the desired set-point. A complete range (0-15 mg/l) of D.O. concentration levels were available through the controller. Simultaneously, the D.O. displayed the D.O. concentration level and recorded these data on a 15-day strip chart.

The responsiveness of the controller was observed to hold D.O. concentration to a maximum variation of ± 2 mg/l from the set-point. In most cases the D.O. concentration variation was ± 1 mg/l with the set-point operating at 3 mg/l.

The D.O. probe was the key to dependable operation of the automatic oxygen control system. For this project, the automatic control system performed without a malfunction. Experience with the D.O. probe established standard probe operation and maintenance (O&M) procedures now employed. These O&M procedures do not require extensive operating time, highly qualified personnel, or complicated items of equipment. It was determined that a 2-mil-thick Teflon membrane was found to provide the best results for continuous monitoring of mixed liquor D.O. concentration. The agitator assembly that is located just below the membrane was adjusted so that it was as close as possible, approximately 1.59 mm to 0.79 mm (1/16 in. to 1/32 in.) to the membrane surface. This condition reduced solids growth on the membrane to a non-interfering level. The life of the 2-mil membrane was observed to be a minimum of 3 mo with membrane life greater than 6 mo expected.

The Weston & Stack probe-analyzer was calibrated against a known D.O. concentration of tap water at least once every 2 weeks. In all cases except one, the Weston & Stack D.O. concentration was within ± 0.5 mg/l range of the known D.O. concentration. The one exception was the result of a membrane failure which occurred after approximately three months of operation. The probe-analyzer was not recalibrated when D.O. concentration variation was within ± 0.5 mg/l.

The D.O. concentrations of the mixed liquor in all three passes were observed using a Yellow Springs Instrument (YSI) D.O. meter, Model 54. It was concluded that the variation of D.O. concentration in "B" and "C" passes was not of enough significance to warrant installation of automatic D.O. control systems. Most of the variation in demand resulted in "A" pass because of variations in the sewage organic loading, with stable demand observed in "B" and "C" passes. For this reason, the oxygen flow to "B" and "C" passes was adjusted manually.

MONITORING METHODS

The monitoring procedures for the pilot plant program were conducted in two phases. Phase I included three of the seven tests shown in Table 1. Phase II included six of the seven tests shown in Table 1 with two of the

test in Phase I being retested during Phase II.

Phase I

This operation consisted of full coverage sample collection, limited operational controls, and moderate sample analysis. During Phase I samples of the SI after screening final clarifier effluent (FE), mixed liquor (ML) and RAS were collected by Englewood plant operators at 2-hr intervals, 24 hours a day, 7 days a week. These samples were composited and refrigerated each day during the week. Each week a composite sample was collected for Friday, Saturday and Sunday. Samples were not composited proportional to flow.

Limited operating controls were used during Phase I relative to D.O. control, sludge blanket depth, and sludge wasting. No automatic D.O. control was provided because of equipment delivery delays and receipt of faulty equipment. The sludge blanket depth and sludge wasting was accomplished by MAROX personnel during the daylight hours, 5 days a week, with occasional sludge wasting during the weekend. The waste activated sludge (WAS) sample was composited proportional to flow.

Table 2 shows the analyses which were conducted on composite samples during Phase I.

TABLE 2. SAMPLE SOURCE AND LABORATORY TESTING IN PHASE I

Sample Source	Total Suspended Solids (TSS)	Total Volatile Suspended Solids (VSS)	Biochemical Oxygen Demand 5 day (BOD_5)
Secondary Influent (SI)	X		X
Final Clarifier Effluent (FE)	X		X
Mixed Liquor (ML)	X	X	
Return Activated Sludge (RAS)	X	X	
Waste Activated Sludge (WAS)	X	X	

The BOD_5 tests using two dilutions were conducted by Englewood lab personnel 5 days a week. The TSS and VSS tests were conducted by MAROX personnel 5 days a week. All tests were conducted in accordance with Standard Methods procedures.[4] At 8-hr intervals the Englewood operators measured the D.O. concentration of the ML in each pass and the D.O. concentration of the supernatant in the final clarifier; recorded the oxygen pressure, flow, and temperature readings; and conducted a settling test with a 1000-ml graduated cylinder from which the calculation of the sludge volume index (SVI) was accomplished. The average 30-min settling result and composite MLSS concentration were used in the SVI calculation.

Phase II

Monitoring procedures were expanded to include increases in the sampling analysis from 5 to 7 days a week and the number of dilutions for BOD_5 tests from 2 to 3 dilutions. The laboratory analyses were expanded to include chemical oxygen demand (COD), total Kjeldahl nitrogen (TKN), ammonia nitrogen (NH_3-N), nitrite nitrogen (NO_2-N), nitrate nitrogen (NO_3-N), turbidity, and pH. Field operations were changed to incorporate a plant operational control scheme consisting of centrifuge tests, depth of blanket determinations, final effluent turbidity measurements, and mixed-liquor settling characteristics using a Mallory settleometer. Automatic D.O. control was operational in the first pass of the oxygenation basin, and process control was improved by increasing sludge wasting from 5 to 7 days a week with samples composited proportional to flow. Because of expanded operational controls during Phase II, samples of SI, FE, ML and RAS were collected at 4-hr intervals compared to the 2-hr intervals of Phase I. These samples were not flow proportioned, but were composited and refrigerated each day of the week.

Laboratory tests were conducted by contracted experienced laboratory personnel using MAROX laboratory equipment, and by Englewood laboratory personnel. TKN, NH_3-N, NO_2-N, and NO_3-N tests were conducted on weekly samples by an independent laboratory. Table 3 indicates the sample sources and laboratory analyses that were performed for Phase II testing.

TABLE 3. SAMPLE SOURCE AND LABORATORY TESTING IN PHASE II

Sample Source	TSS	VSS	BOD_5	COD	TKN	NO_2-N	NO_3-N	NH_4-N	pH	Turb
SI	X	X	X	X	X	X	X	X	X	
FE	X		X	X	X	X	X	X	X	X
ML	X	X								
RAS	X	X								
WAS	X	X		X*						

* Intermittent

All analyses were conducted in accordance with Standard Methods with the following modifications.

1. The TSS and VSS analyses were conducted using a Reeve Angel grade 934AH base mat and a Pyrex fiber glass super mat in a Coors Gooch crucible No. 27007-4. Drying time was one hour at 103° ± 1.0° C. Incineration time was 15 was at 600° C.

2. The COD analyses were conducted using 50 ml of sample, 25 ml of 0.250 N potassium dichromate, 75 ml of sulfuric acid with silver sulfate, and 0.250 N ferrous ammonium sulfate as the titrant. The potassium dichromate used was a primary standard dried for 3-hr at 103° C ± 1.0° C A Friedrich's condenser reflux apparatus was used with the reflux time equal to 2-hr.

3. The turbidity test was conducted using a Hack turbidity meter, Model 5100 A. Hack Formazn turbidity standards were used to calibrate the meter. The turbidity was measured in Formazin Turbidity Units (FTU) and was recorded as equivalent Jackson Turbidity Units (JTU).

Each day Englewood plant operators performed centrifuge "spin" tests on the mixed liquor and return activated sludge, depth of sludge blanket in the clarifier and clarifier effluent turbidity measurements, and mixed liquor settling characteristics using a Mallory settleometer. The results were used to identify the relative position of the total mass of sludge in the system, monitor variations in effluent quality, and determine changes in sludge settling characteristics. The procedures used to conduct these tests are described by West[5] and Hegg.[6]

The other operational tests described in Phase I were conducted by Englewood operators twice each day. These included D.O. concentration of the ML in each pass, D.O. concentration of the supernatant in the clarifier, and settling test to indicate SVI. The frequency of these tests were reduced from three times in Phase I to two times in Phase II because of the increased operational load that was placed upon the operators.

The test monitoring for oxygen pressure, flow, and temperature readings remained the same as Phase I. In both phases, average values for these readings were used to calculate the quantity of oxygen supplied to the system at a standard temperature of 20° C (68° F) and a standard pressure of 760 mm (30 in.).

SECTION 5

DISCUSSION OF RESULTS

PHASE I

The Phase-I experimental program consisted of the first three test conditions identified in Table 1. Table 4 shows the average monitoring results, and Table 5 shows the average operating results recorded during Phase I. Most data listed in Table 4 are self-explanatory. The items not described earlier are raw sewage BOD_5 and TSS concentrations. These values are Englewood laboratory results for the raw sewage entering the treatment facility.

TABLE 4. MONITORING RESULTS - PHASE I

	Run 1	Run 2	Run 3
Operating conditions			
Average SI flow, l/min (gpm)	37.8 (10)	37.8 (10)	56.7 (15)
Flow pattern	Constant	Diurnal	Constant
Run length, days	47	25	54
Sewage temperature, ^{o} C	22	20	13
BOD_5 data			
Raw sewage, mg/l	183	179	245
SI, mg/l	131	140	147
Primary removal, %	18.4	21.8	40.0
FE, mg/l	10	8	16
Secondary removal, %	92.4	94.3	89.8
Primary plus secondary removal, %	94.5	95.5	93.5
Oxygen supplied, mass O_2/mass BOD_R	2.30	2.60	1.17
TSS data			
Raw sewage, mg/l	197	153	157
SI, mg/l	187	179	124
Primary removal, %	5.1	17*	21
FE, mg/l	19	21	26
Secondary removal, %	84.4	87.7	78.8
Primary plus secondary removal, %	90.4	86.3	83.4

* This indicates an increase of 17%.

TABLE 5. OPERATING RESULTS - PHASE I

	Run 1	Run 2	Run 3
Operating conditions			
Average SI flow, l/min (gpm)	37.8 (10)	37.8 (10)	56.7 (15)
Flow pattern	Constant	Diurnal	Constant
Run length, days	47	25	54
Volumetric loading, kg $BOD_5/m^3/day$ (lb $BOD_5/1000\ ft^3/day$)	0.96 (60)	1.03 (64)	1.71 (107)
Aerator detention time (Q), hr	3.3	3.3	2.2
Sludge detention time-aerator (Q+R), hr	2.4	2.3	1.6
RAS* flow, l/min (gpm)	14.8 (3.9)	16.7 (4.4)	20.8 (5.5)
R/Q ratio, %	39.4	43.6	36.4
Sludge detention time-clarifier (Q+R), hr	2.0	3.9	2.8
Clarifier mass loading, kg TSS/day/m^3 (lb TSS/day/ft^3)	161 (33.0)	220 (45.0)	152 (31.2)
Overflow rate**, $m^3/day/m^2$ (gpd/ft^2)	15.2 (374)	15.2 (374)	22.7 (561)
Overflow rate+, $m^3/day/m^2$ (gpd/ft^2)	16.7 (411)	16.7 (411)	25.1 (617)
Sludge age++, days	13.9	13.0	3.3
Mean cell residence time, days	25.6	27.8	7.1
Sludge synthesis, mass excess VSS/ mass BOD_5 removed	0.44	0.57	0.77
F/M ratio, mass BOD_5/mass MLSS/day	0.18	0.15	0.46
Oxygen data			
Pass-A mixed liquor DO, mg/l	9.3	9.1	3.7
Pass-B mixed liquor DO, mg/l	9.5	10.1	3.6
Pass-C mixed liquor DO, mg/l	8.0	10.5	4.3
Final effluent DO, mg/l	4.1	5.7	1.8
Solids data			
MLSS, mg/l	7,488	10,206	4,877
MLVSS, mg/l	5,390	7,111	3,870
MLVSS/MLSS ratio, %	72.0	69.7	79.4
RAS TSS, mg/l	26,755	24,448	14,139
RAS VSS, mg/l	19,325	17,234	10,951
RASVSS/RASTSS ratio, %	72.2	70.5	77.5
SVI, ml/gm	65.9	67.7	81.5

*RAS = return activated sludge
**Includes clarifier centerwell.
+Excludes clarifier centerwell.
++Based on TSS in secondary influent.

For the calculated values of Table 5, the following methods were used:

1. FM = Mass BOD_5 applied per day/mass of MLVSS.

2. Volumetric loadings = Mass BOD_5 applied per day/unit volume of oxygenation basin.

3. Aerator detention time = Aerator detention time based on SI.

4. Sludge detention time aerator = Aerator detention time based on SI + RAS.

5. SVI = Daily average 1000 ml graduated cylinder 30 minute settling value/by composite MLSS value.

6. R/Q ratio = RAS flow/by the SI flow.

7. Sludge detention time clarifier = Mass of sludge in clarifier/by mass of sludge recycled per hour. NOTE: The sludge volume in the clarifier was determined using the sludge blanket depth values. The sludge concentration in the clarifier was assumed to be equal to the average of the return sludge and mixed liquor TSS concentration.[7]

8. Clarifier mass loading = Mass of solids applied to clarifier per day (includes SI plus RAS)/by the clarifier surface area.

9. Total overflow rate = SI flow divided by the clarifier surface area.

10. Overflow rate = SI flow/by the clarifier surface area excluding the centerwell.

11. Sludge synthesis = Mass of volatile sludge wasted per day (including sludge over clarifier weir)/by mass of BOD_5 removed day.

12. Sludge age = Mass of volatile solids in oxygenation basin/by mass of volatile solids wasted per day (includes volatile solids lost over clarifier weir).

13. Mean cell residence time = Mass of volatile solids in aerator and clarifier/by mass of volatile solids wasted per day (includes volatile solids lost over clarifier weir).

Phase I process removal efficiencies were satisfactory despite limited operational controls and low primary clarifier TSS removal efficiencies. MAROX secondary BOD_5 removal efficiencies ranged from 89.8 percent to 94.3 percent with TSS removal efficiencies from 78.8 percent to 87.7 percent. Primary treatment BOD_5 removal efficiencies were in an expected range of 21.8 percent to 40 percent. Primary treatment TSS removal efficiencies were much lower than expected, ranging from a negative 17 percent to a positive 21

percent. The low primary TSS removal efficiencies may be attributed to higher than desired TSS concentrations of various Englewood in-plant recycle flows, namely anaerobic digester supernatant, that may have adversely affected the final effluent quality.

Final effluent during Phase I had BOD_5 concentrations less than 17 mg/l and TSS concentrations less than 27 mg/l. The slightly higher TSS values apparently were comprised of inert suspended solids not captured in the final clarifier. The final effluent probably contained more inert suspended solids than normal because of two reasons: (1) a long sludge detention time in the clarifier (2.0 to 3.9 hr) which contributed to degradation of mixed liquor quality, and (2) a high concentration of anaerobic digester supernatant solids in the SI flow. Degradation of mixed liquor quality was demonstrated by its poorer ability to capture solids in the clarifier. The long sludge detention time in the clarifier was caused by the lower return sludge flow rate than was established in Phase II. The excellent settling characteristics of the mixed liquor (SVI ranged from 65.9 ml/gm to 81.5 ml/gm) did not result in a clean-sweeping sludge. If the R/Q ratio had been increased, the lower sludge detention time in the clarifier might have improved the mixed liquor quality.

High concentrations of anaerobic digester supernatant solids existed in the SI. Primary clarifier TSS removal efficiencies were much less than normal indicating high concentrations of recycle TSS in the primary clarifier. These digesters supernatant solids may have contributed to the apparent inert TSS concentrations in the final effluent.

The average D.O. concentration of the ML in each pass of the oxygenation basin was higher than desired during the first two Phase I tests. During the last part of Phase I, the average D.O. concentration of the ML was in the desired range. During all periods of Phase I operation manual D.O. control was provided.

In parts of Phase I operation, the mass of oxygen supplied per mass of BOD_5 removed was larger than expected, especially during those periods when the mixed liquor D.O. was higher than desired.

The high oxygen requirements for the first two Phase I tests were caused by higher than desired mixed liquor D.O. concentrations, and high autodigestion oxygen demand as a result of long mean cell residence times.

Phase I pilot plant operation demonstrated that the MAROX system can be satisfactorily operated at high MLVSS concentrations, high return sludge TSS concentrations, and long mean cell residence times or sludge ages.

PHASE II

Table 6 shows the average monitoring results, and Table 7 shows the average operating results during Phase II. The experimental program for Phase II included the last four test conditions cited in Table 1 plus reruns of test conditions Nos. 1 and 3.

TABLE 6. MONITORING RESULTS - PHASE II

	Run 1	Run 2	Run 3	Run 4	Run 5	Run 6
Operating conditions						
Average SI flow, l/min (gpm)	37.8 (10)	56.7 (15)	75.8 (20)	94.5 (25)	113.5 (30)	132.5 (35)
Flow pattern	Constant	Constant	Diurnal	Constant	Diurnal	Constant
Run length, days	30	31	31	23	30	31
Sewage temperature, ^{o} C	18.1	19.5	21.5	14.6	19.0	16.8
SI pH	6.8	6.9	6.9	7.0	6.9	7.0
FE pH	6.7	6.9	6.9	6.9	6.9	6.9
FE turbidity, JTU	4.7	4.6	5.1	5.6	4.4	6.3
BOD_5 and COD data						
Raw sewage BOD_5, mg/l	282	292	289	191	297	262
SI BOD_5, mg/l	212	210	208	166	179	159
Primary BOD_5 removal, %	24.8	28.1	17.6	13.1	39.7	39.3
FE BOD_5, mg/l	15	16	16	14	16	18
Secondary BOD_5 removal, %	92.9	92.4	92.3	91.6	91.1	88.7
Primary plus secondary BOD_5 removal, %	94.7	94.5	94.5	92.7	94.6	93.1
Oxygen supplied, mass O_2/mass BOD_R	1.49	1.32	1.08	1.17	0.92	1.08
SI COD, mg/l	396	345	321	332	259	305
FE COD, mg/l	48	48	61	95	57	74
COD removal, %	87.9	85.5	80.1	70.9	77.6	75.7
TSS data						
Raw sewage, mg/l	205	202	175	131	233	200
SI, mg/l	134	119	115	110	85	85
Primary removal, %	34.6	41.1	34.3	16.0	63.5	57.5
FE, mg/l	12	14	15	13	11	15
Secondary removal, %	91.4	86.7	85.6	87.7	85.8	81.4
Primary plus secondary removal, %	94.1	93.1	91.4	90.1	95.3	92.5
Nitrogen data						
SITKN, mg/l	31	27	27	36	29	26
FE TKN, mg/l	20	13	12	28	23	21
TKN removal, %	34.2	51.9	55.6	27.8	20.7	19.2
SI NH_3-N, mg/l	18	19	18	29	21	18
FE NH_3-N, mg/l	16	16	15	22	19	15
NH_3-N removal, %	10.6	15.8	16.7	24.1	9.5	16.7
SI NO_2-N + NO_3-N, mg/l	0.25	0.1	0.1	0.1	0.5	0.1
FE NO_2-N + NO_3-N, mg/l	0.25	0.15	0.15	0.15	1.3	0.16

TABLE 7. OPERATING RESULTS - PHASE II

	Run 1	Run 2	Run 3	Run 4	Run 5	Run 6
Operating conditions						
Average SI flow, l/min (gpm)	37.8 (10)	56.7 (15)	75.8 (20)	94.5 (25)	113.5 (30)	132.5 (35)
Flow pattern	Constant	Constant	Diurnal	Constant	Diurnal	Constant
Run length, days	30	31	31	23	30	31
Volumetric loading, kg $BOD_5/m^3/day$ (lb $BOD_5/1000\ ft^3/day$)	1.54 (96)	2.31 (144)	3.04 (190)	3.03 (189)	3.91 (244)	4.07 (254)
Aerator detention time (Q), hr	3.3	2.2	1.6	1.3	1.1	0.94
Sludge detention time-aerator (Q+R), hr	2.12	1.40	1.10	0.83	0.70	0.62
RAS* flow, l/min (gpm)	20.6 (5.45)	30.2 (8.0)	37.8 (10.0)	56.7 (15.0)	64.6 (17.1)	69.2 (18.3)
R/Q ratio, %	54.5	53.3	50.1	60.0	57.0	52.3
Sludge detention time-clarifier (Q+R), hr	3.3	2.5	1.4	1.0	1.4	1.5
Clarifier mass loading, kg TSS/day/m^3 (lb TSS/day/ft^3)	112 (23)	181 (37)	181 (37)	156 (32)	171 (35)	161 (33)
Overflow rate**, $m^3/day/m^2$ (gpd/ft^2)	15.2 (374)	22.9 (561)	30.5 (748)	19.1 (468)	15.2 (561)	26.7 (655)
Overflow rate+, $m^3/day/m^2$ (gpd/ft^2)	16.7 (411)	25.1 (617)	33.5 (822)	20.9 (514)	25.1 (617)	29.3 (720)
Sludge age++, days	4.5	3.2	2.6	2.0	2.2	1.5
Mean cell residence time, days	11.2	8.3	5.8	4.1	6.2	4.9
Sludge synthesis, mass excess VSS/ mass BOD_5 removed	0.61	0.60	0.46	0.74	0.50	0.61
F/M ratio, mass BOD_5/mass MLSS	0.39	0.57	0.95	0.77	1.02	1.23
Oxygen data						
Pass A mixed liquor DO, mg/l	3.0	3.5	1.7	3.8	3.7	4.4
Pass B mixed liquor DO, mg/l	2.7	4.2	2.3	4.0	5.2	4.5
Pass C mixed liquor DO, mg/l	3.1	4.5	3.9	4.8	4.4	6.1
Final effluent DO, mg/l	1.2	1.2	1.0	0.6	1.3	1.3
Solids data						
MLSS, mg/l	4,750	5,089	3,962	5,122	4,743	4,028
MLVSS, mg/l	3,952	4,044	3,361	4,008	3,886	3,436
MLVSS/MLSS ratio, %	83.2	79.5	84.8	78.4	81.9	85.3
RAS TSS, mg/l	13,103	13,131	10,850	11,583	12,221	10,487
RAS VSS, mg/l	10,842	10,390	9,080	9,078	9,820	8,695
RASVSS/RASTSS ratio, %	82.7	79.1	83.7	78.4	80.4	82.9
SVI, ml/gm	204	184	74.5	71.5	67.8	72.4

*RAS = return activated sludge
**Includes clarifier centerwell.
+Excludes clarifier centerwell.
++Based on TSS in secondary influent.

No extremely high or low values for the MLVSS concentration, R/Q ratio, return activated sludge TSS concentration, mean cell residence time or sludge age were recorded. All six periods of operation during Phase II demonstrated excellent BOD_5 and TSS removals. Final effluent BOD_5 and TSS concentrations ranged from 14 to 18 mg/l, and from 11 to 15 mg/l, respectively.

Primary treatment BOD_5 and TSS removal efficiencies were acceptable, ranging from 13.1 percent to 39.7 percent, and from 16 percent to 63.5 percent, respectively. Phase II primary TSS removal efficiency as low as 16 percent occurred one time only at 95 l/min (25 gpm) flow to the system. The low primary TSS removal efficiency of 16 percent is attributed to anaerobic digester supernatant recycle flow. However, during this period, the final effluent TSS concentration was much improved over the Phase I overall final effluent TSS concentration. Two reasons are given as contributing to the better effluent quality during this period: (1) operation at higher R/Q ratio, and (2) lower sludge detention time in the clarifier. The R/Q ratio and clarifier sludge detention time were 50 percent and one hour, respectively. During Phase I, these same parameters were approximately 40 percent and 3 hr, respectively.

It should be noted that during the 95 l/min (25 gpm) flow condition, the secondary influent TSS concentration was lower than the overall values for Phase I. The lower secondary influent TSS concentration may have contributed to the final effluent TSS concentration of 13 mg/l. In all but two periods of Phase II, (37.8 l/min (10 gpm) and 56.7 l/min (15 gpm)) the sludge detention time in the clarifier was less than 1.4 hr. In these two periods, however, the sludge detention time in the clarifier was 3.29 hr and 2.5 hr, with SVI values of 204 ml/gm and 184 ml, respectively. Although the settling characteristics were poorer than desired, the final effluent quality did not deteriorate (12 and 14 mg/l, respectively). As mentioned under Phase I discussion, the sludge in these two periods was a clean-sweeping sludge even though its settling rate was not as good. The sludge settling characteristics of the other four periods were excellent with SVI ranging from 67.8 ml/gm to 74.5 ml/gm.

The D.O. concentration of the mixed liquor in each pass was within the satisfactory range of 1 mg/l to 4 mg/l.

During Phase II, the automatic oxygen controller was in operation and performed very effectively. The mass of oxygen supplied per mass of BOD_5 removed varied from 0.92 to 1.49. The higher value was obtained at the lowest hydraulic flow tested showing a greater oxygen requirement for autodigestion. At higher SI flow, the mass of oxygen supplied per mass of BOD_5 removed was very close to 1.0.

Results of the total nitrogen series analyses indicate that no significant nitrification occurred during Phase II. Limited TKN and NH_3-N removals were observed. Results of the COD analyses indicated that the relationship of final effluent COD to final effluent BOD_5 was very erratic. The final effluent COD concentration ranged from 48 to 94 mg/l, whereas the final effluent BOD_5 concentration ranged from 14 to 18 mg/l. The final effluent turbidity test results were low and consistently ranged from 4.4 to 6.3 JTU.

No significant decrease in pH occurred between the SI and FE during Phase II. The average pH of the FE during three periods of operation was equal to the average pH of the SI. During the other three periods of operation, the average pH of the FE was 0.1 unit less than the average pH of the SI. It is apparent that most of the CO_2 produced during the biological reaction was removed from the system in a form which did not significantly decrease the pH of the final effluent.

SECTION 6

ANALYSIS OF RESULTS

Included in this section is an evaluation of both phases of operation and an evaluation of the special testing completed. The evaluation is made with special reference to system performance analysis, oxygen utilization analysis, sludge production analysis, and sludge settling analysis.

SYSTEM PERFORMANCE ANALYSIS

The treatment performance of the MAROX system for both Phase I and Phase II operation is graphically depicted in Figures 5, 6, 7, and 8. Figure 5 shows the BOD_5 concentration of the raw sewage, secondary influent, and final effluent for all test conditions. The average final effluent BOD_5 concentration for each test condition never exceeded 20 mg/l. Figure 6 shows the BOD_5 removal for both the primary treatment and the primary plus MAROX secondary treatment. Primary treatment removals showed wide variations with primary plus secondary removals always exceeding 90 percent.

Figure 7 shows the TSS concentration of the raw sewage, SI and FE for all test conditions. The TSS concentration of the SI exceeded the TSS concentration of the raw sewage during one test condition in Phase I. The TSS concentration of the FE never exceeded 30 mg/l during Phase 1, nor 20 mg/l during Phase II. Figure 8 shows the primary treatment TSS removal efficiencies for all tests. Primary removal efficiencies showed wide variations even including a negative TSS removal. Primary plus secondary TSS removal efficiencies were greater than 80 percent during Phase I and greater than 90 percent during Phase II.

During the study, the pilot facility was operated under a wide range of loading conditions. At the higher loading conditions, no significant decrease in process removal efficiency was noted. Figure 9 depicts the BOD_5 substrate removal rate related to the F/M ratio. The graph was developed using weekly average test results for all test conditions. The straight-line relationship of the graph illustrates that no degradation in process BOD_5 removal efficiency existed even at the higher F/M ratios. The slope of the line indicated that on the average for all tests, 90.2 percent of the BOD_5 applied to the system was removed.

OXYGEN UTILIZATION ANALYSIS

The oxygen utilization efficiency in closed-tank systems is calculated by comparing the quantity of oxygen supplied to the quantity of oxygen in the off-gas. The open tank system admittedly cannot readily quantify tank off-gas oxygen mass. For this reason, three alternative procedures for determining

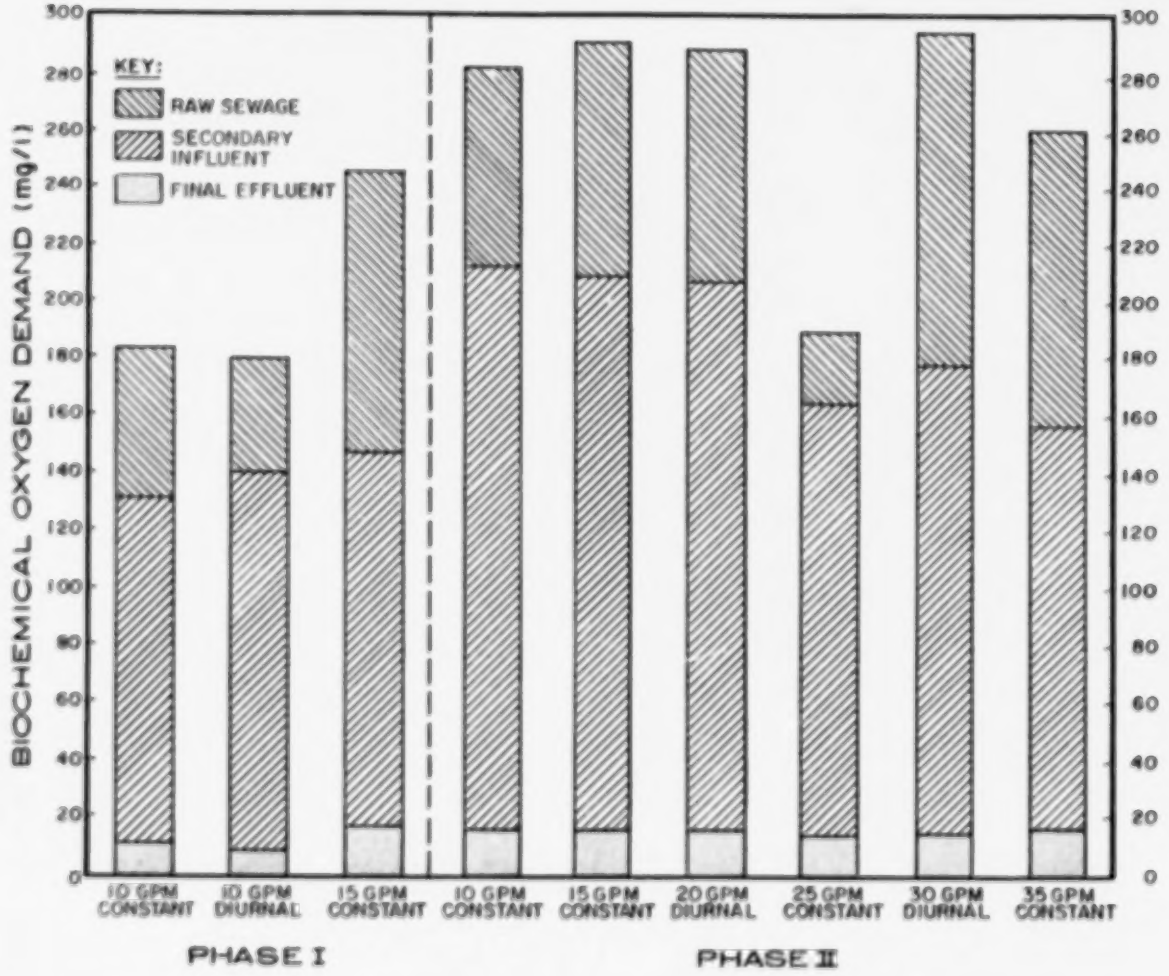

Figure 5. BOD concentration in raw sewage, secondary influent, and final effluent.

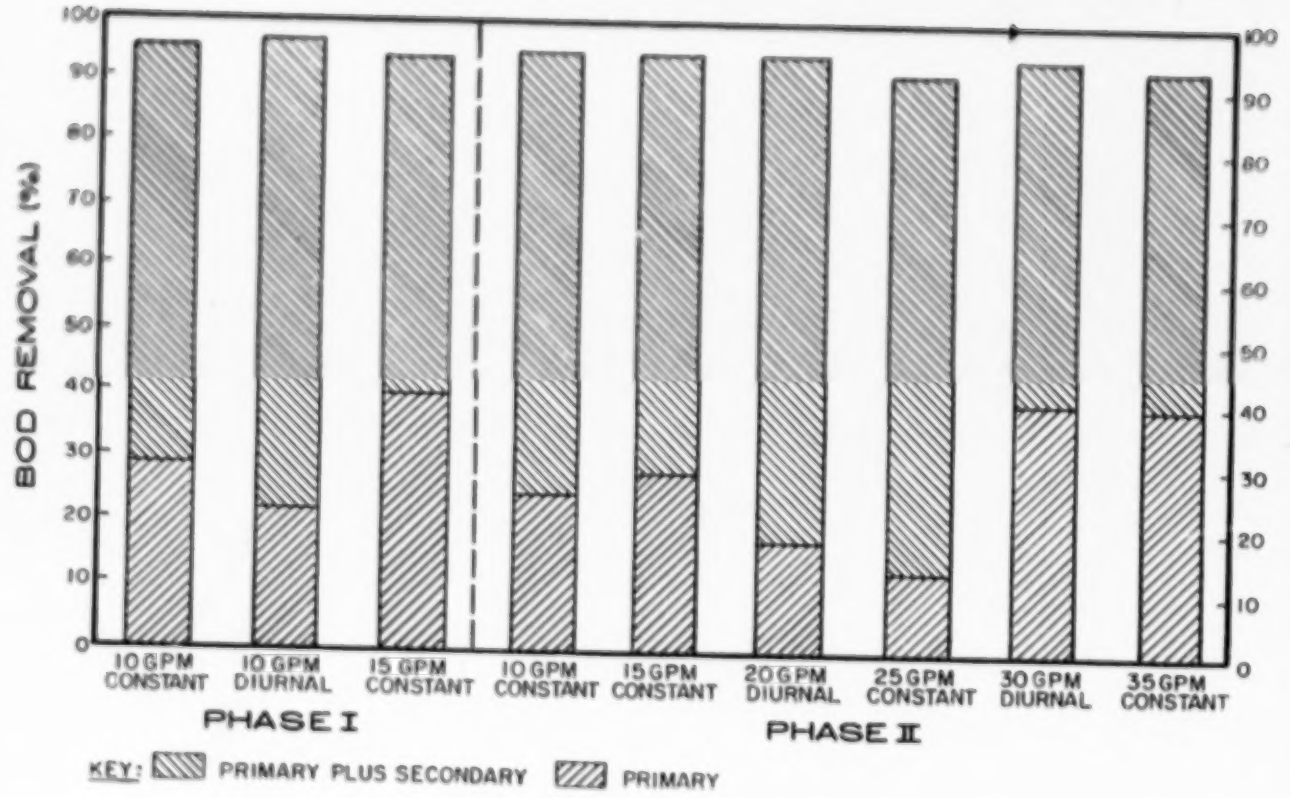

Figure 6. BOD removal performance.

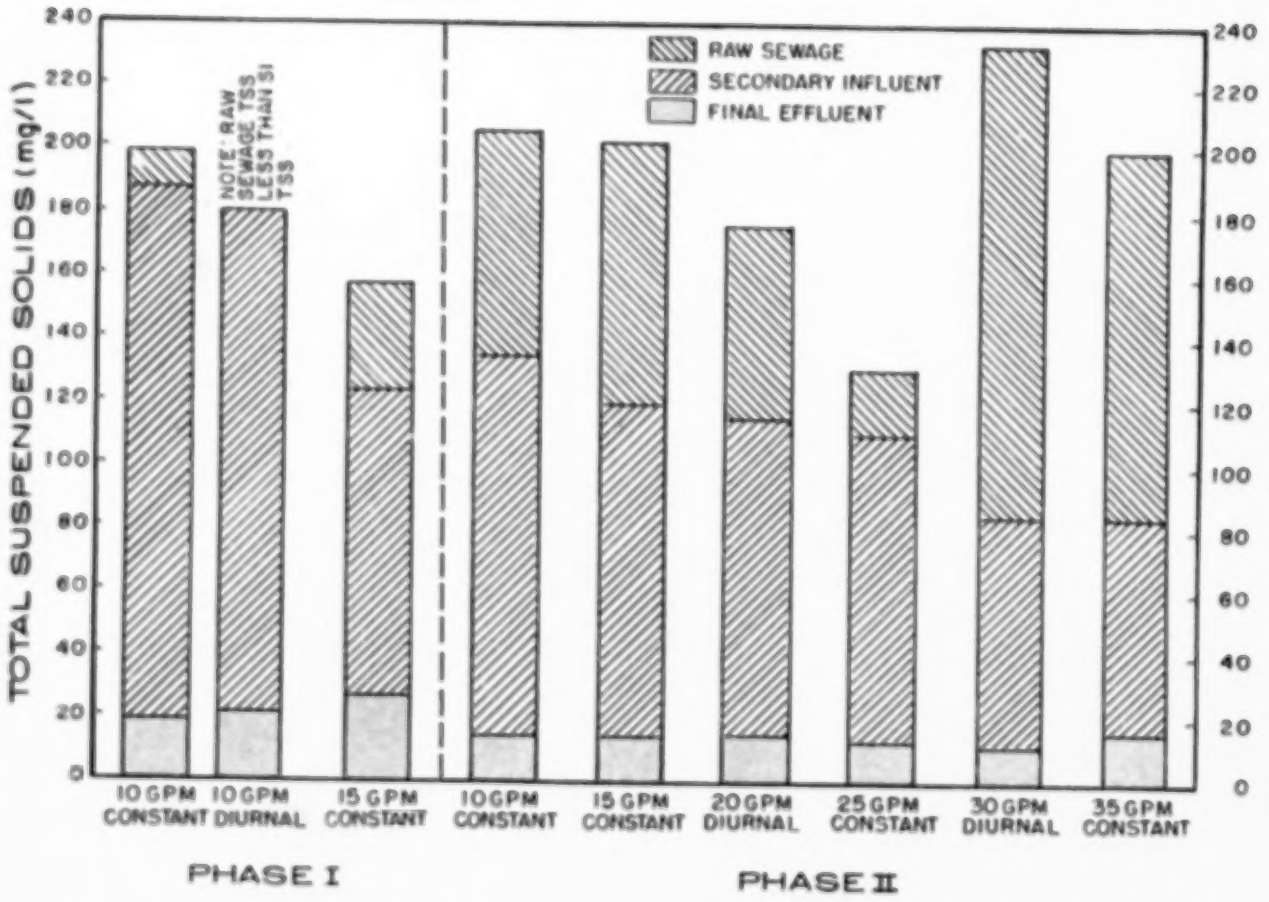

Figure 7. TSS concentration in raw sewage, secondary influent, and final effluent.

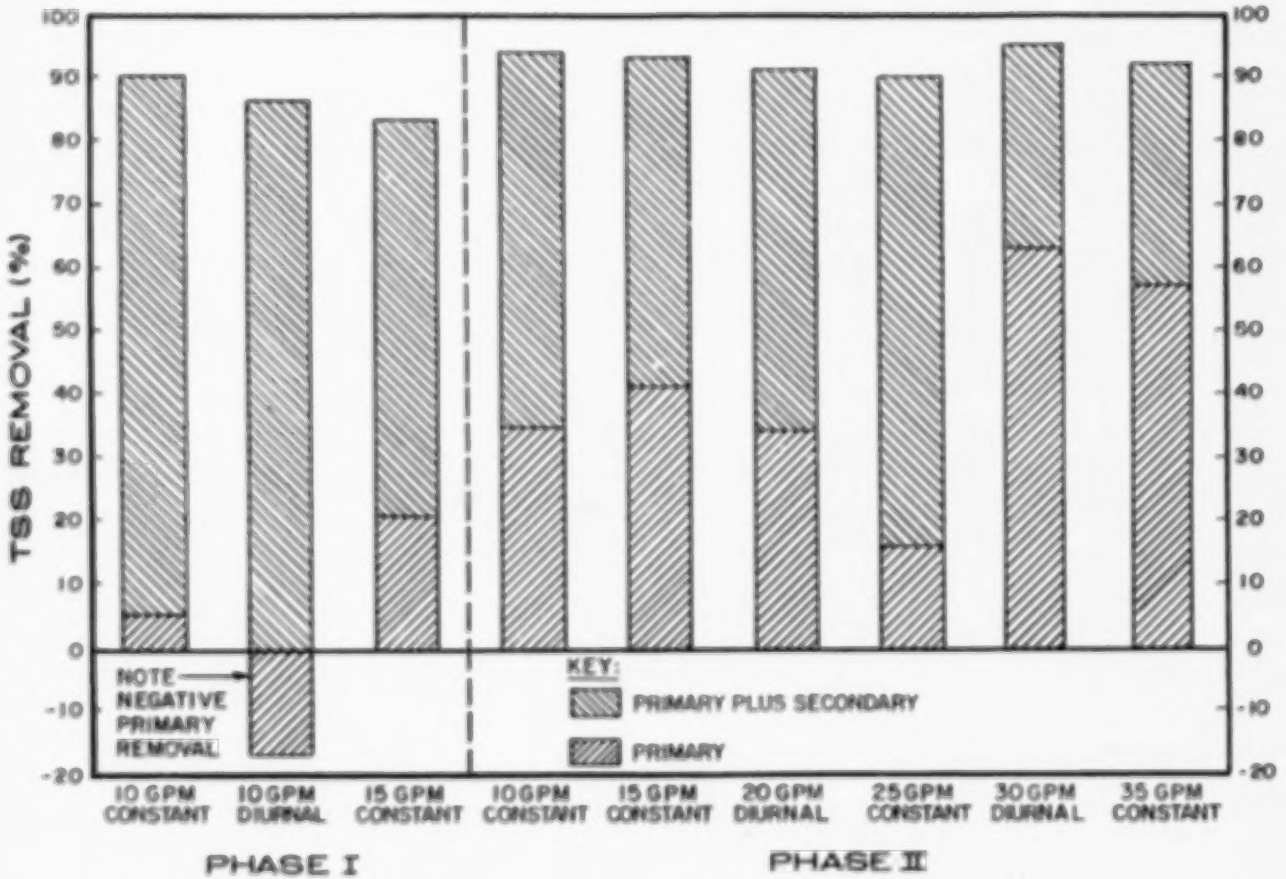

Figure 8. TSS removal performance.

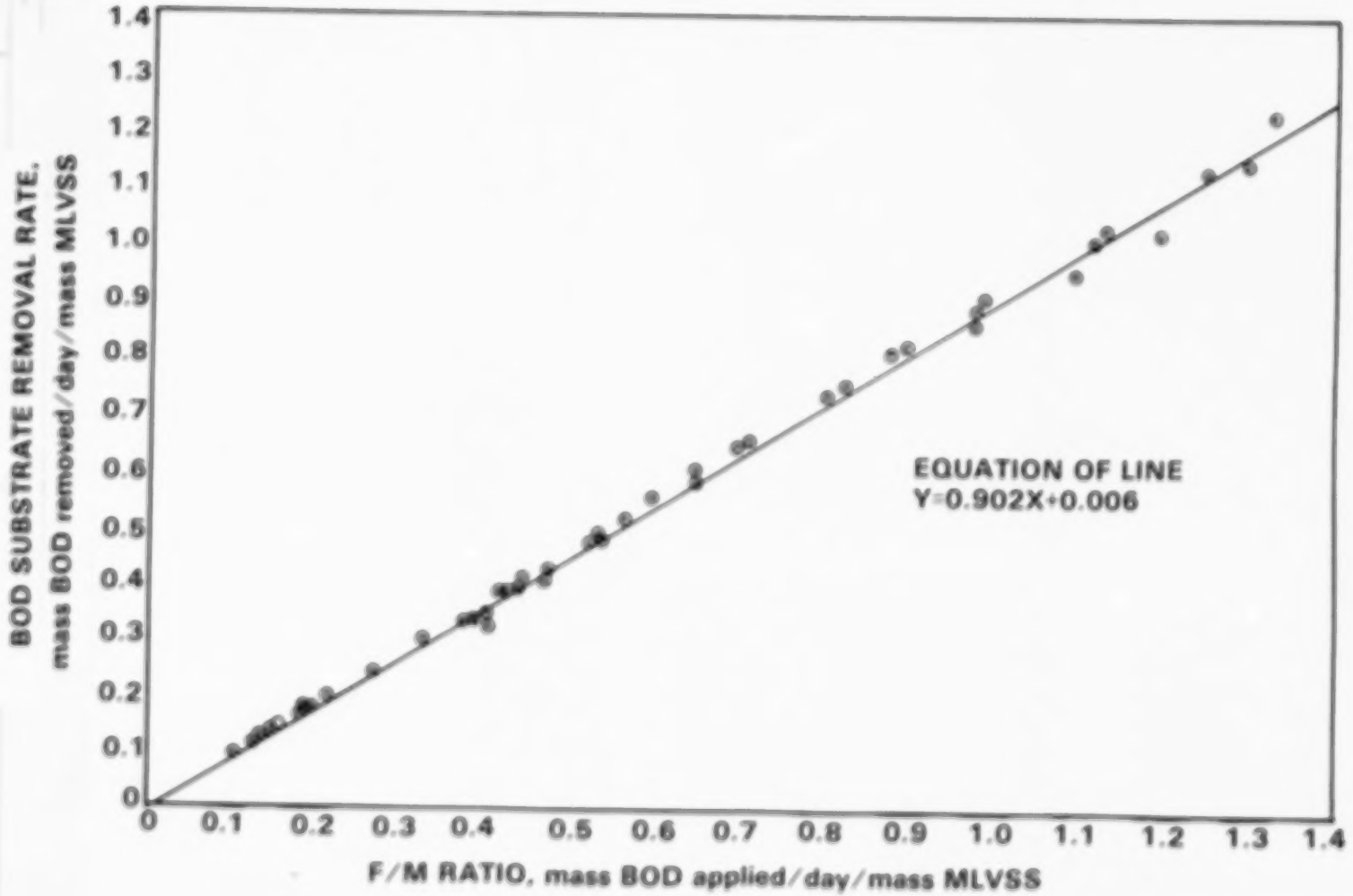

Figure 9. BOD substrate removal rate versus F/M ratio.

Oxygen utilization efficiencies were investigated. These Alternatives compared the mass of oxygen supplied with the mass of BOD_5 removed, the mass of oxygen supplied with the mass of COD removed, and the mass of oxygen supplied with the mass of oxygen contained in the off-gas analyzed during special test periods when the top of the aeration reactor was capped. Each alternative was analyzed with respect to identifying the best procedure for determining oxygen utilization efficiency with the open tank system.

Of the three alternatives, the off-gas method is more direct and more accurate as shown in the following discussion. In order to capture off-gas, a special cover of 6 mil thick polyethylene sheet was installed on the top of the aeration basin. Gas leak checks were made to insure that a tight seal existed between the polyethylene cover and the top of the aeration basin. One vent line was tapped into the space above the surface of the ML. In all cases, the gas pressure in the ullage space was maintained at a pressure less than 6.35 mm (1/4 in.) of water, so that a negligible quantity of oxygen could be re-introduced into the ML. After the top of the oxygenation tank was capped, the system was operated for one full day to sufficiently purge the system before the off-gas was analyzed. Sufficient data was collected then for the determination of oxygen utilization efficiencies. Data collection continued for 4 to 5 days. During the first off-gas analysis test, the off-gas flow was determined periodically throughout the day for selected time periods to calculate oxygen utilization efficiencies. During the second off-gas analysis test, the off-gas flow was measured and totalized using a positive-displacement gas meter. The daily average flow was determined using the recorded, cumulative meter reading once per day. The oxygen utilization efficiency was calculated on a daily basis. During both off-gas analysis tests, the off-gas oxygen concentration was determined using a portable gas test meter with the precision of the portable gas test meter checked against independent laboratory results. The mass of oxygen supplied to the system was determined as described earlier in this report. The results of the off-gas analysis tests are shown in Table 8. Results show that the MAROX diffuser achieved better than 90 percent overall oxygen transfer. The average oxygen usage efficiency for test No. 1 and test No. 2 were 91.5 percent and 92.5 percent, respectively.

The second alternative for determining oxygen utilization efficiency consisted of comparing the mass of oxygen supplied with the mass of BOD_5 removed. This is an indirect method for determining the oxygen utilization efficiency where mass of oxygen utilized was assumed to approximate the mass of BOD_5 removed. This assumption is invalid because oxygen is not only utilized to remove BOD_5, but is also utilized for autodigestion with the quantity of oxygen required for autodigesting being greater at lower F/M ratios. Figure 10 illustrates this fact by comparing the BOD_5 specific oxygen requirement (mass of oxygen supplied per mass of BOD_5 removed) with the F/M ratio (mass of BOD_5 applied per day per mass of MLVSS). Weekly average test results graphically illustrated that additional oxygen is required per mass of BOD_5 removed at the lower F/M ratios.

TABLE 8. OXYGEN UTILIZATION EFFICIENCIES USING OFF-GAS ANALYSIS

Test No.	Off-gas Flow Rate (Std. m^3/min)*	Off-gas O_2 conc (%)	Oxygen Lost (kg)+	Oxygen Supplied (kg)+	Oxygen Utilized (%)
		Test Period No. 1, June 10-14, 1974. Flow Condition: 132.5 l/min (35 gpm)			
1	.00158	33.0	.0122	.2596	95.3
2	.00238	33.9	.0178	.2448	92.7
3	.00282	44.9	.0236	.2236	89.4
4	.00303	44.9	.0236	.2076	88.6
Average	.00245	39.2	.0193	.2339	91.5
		Test Period No. 2, November 5-10, 1974. Flow Condition: 37.8 l/min (10 gpm)			
1	.00164	56.5	.0167	.159	89.5
2	.00114	55.3	.0112	.144	92.2
3	.00131	57.5	.0284	.342	91.7
4	.00063	60.6	.0054	.171	96.7
Average	.00118	57.5	.0154	.204	92.5

* Std. m^3/min x 35.31 = Scfm
\+ kg x 2.205 = lb

There was no practical way of determining the quantity of oxygen required for autodigestion, although an indication of this was determined by comparing the MLVSS specific oxygen requirement to the BOD substrate removal rate as shown in Figure 11. The weekly average test results show that an average of 0.21 kg oxygen/kg of MLVSS was supplied for autodigestion and 0.83 kg oxygen/kg of BOD_5 removed was supplied for substrate removal.

Results of the graph shown in Figure 11 can be used to extrapolate a full scale oxygen requirement value for the specific waste that was treated during the pilot plant operation. Results from Figure 11 cannot be used to determine the quantity of oxygen utilized for autodigestion because the graph was plotted using quantity of oxygen supplied. The quantity of oxygen utilized could not be determined since it is a function of oxygen needed for BOD_5 removed and autodigestion. Attempted measurement of respiration rates in previous test

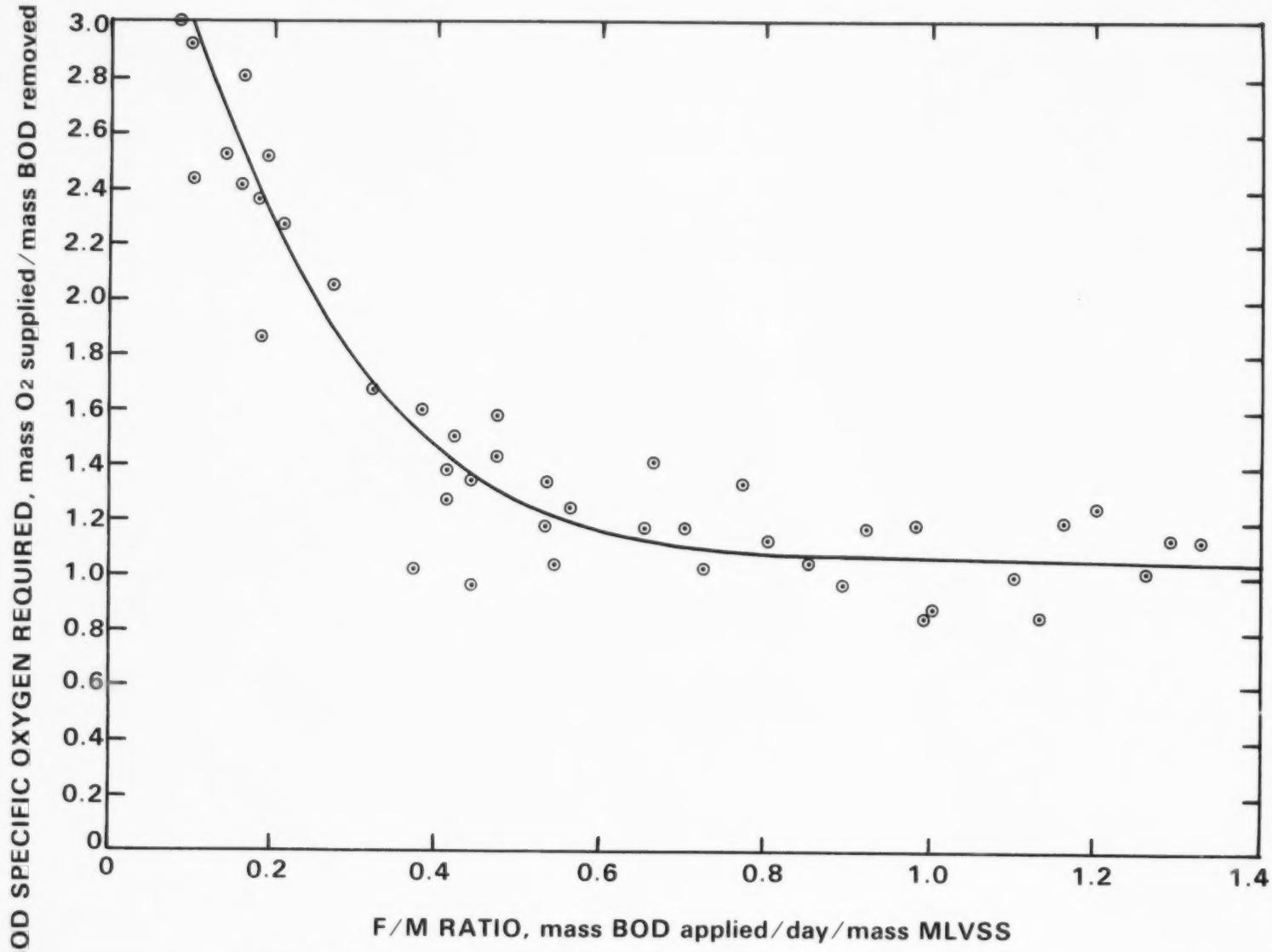

Figure 10. BOD specific oxygen requirement versus F/M ratio.

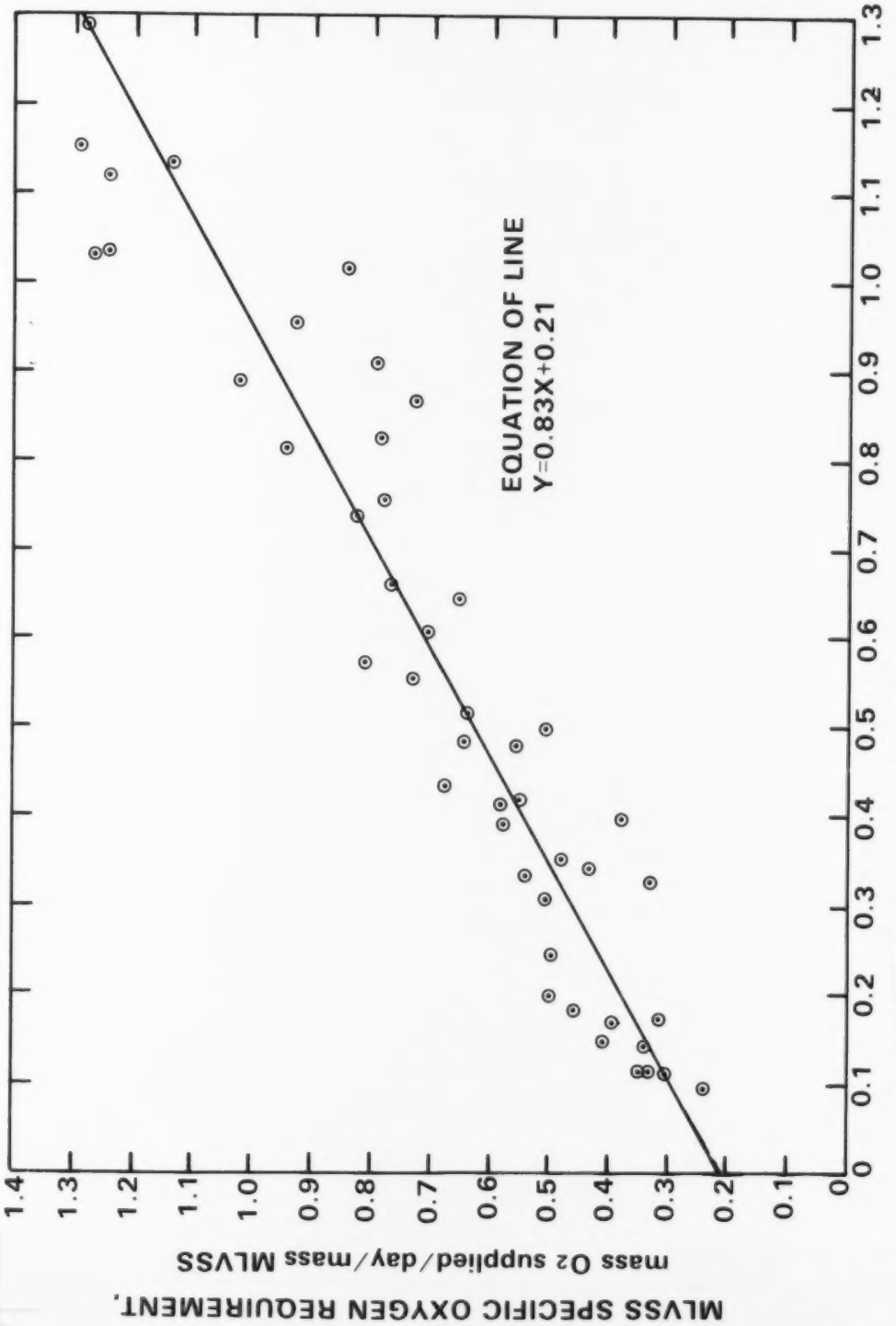

Figure 11. MLVSS specific oxygen requirement versus BOD substrate removal rate.

programs have proved unreliable to evaluate oxygen utilized. In this case, even accurate measurement of oxygen demand would not differentiate between oxygen requirements for BOD_5 requirements and autodigestion. For this reason, the second method is not a recommended alternative for determining specific values for oxygen utilization efficiency.

The third alternative for determining oxygen utilization efficiency consisted of comparing the mass of oxygen supplied with the mass of COD removed. For this comparison it was assumed that 1.0 kg (2.2 lb) of oxygen was utilized for every 1.0 kg (2.2 lb) of COD removed. COD removed was determined using the following equation:

$$COD_R = COD_{SI} - COD_{FE} - COD_{WAS} \pm COD_{Accum}$$

where:

COD_R = COD removed (mass/day)

COD_{SI} = COD of secondary influent (mass/day)

COD_{FE} = COD of final effluent (mass/day)

COD_{WAS} = COD of sludge wasted from the system (mass/day)

COD_{Accum} = COD of the increase or decrease in total sludge inventory of the system (mass/day).

Table 9 shows oxygen utilization efficiencies using this method for five different periods of operation. Results showed that oxygen utilization efficiencies ranged from a negative 115.8 percent to a positive 101.0 percent. Based upon the results of the off-gas analyses as shown in Table 8, greater than 90 percent oxygen utilization was expected. During the time period that test No. 4 of Table 9 was being conducted, the top of the aeration basin was covered and the off-gas was analyzed (Table 8, test No. 2). The oxygen utilization efficiency based on the off-gas analysis averaged 92.5 percent. The oxygen utilization efficiency based on the COD balance for test No. 4 was 17.6 percent, indicating a considerable discrepancy. The oxygen utilization efficiency based on COD balance was questionable due to the change in the sludge inventory during the controlled testing period. For example, a 1.36 kg (3 lb) error in the MLVSS inventory extrapolates to a 10 percent error in the oxygen utilization efficiency. This MLVSS inventory error is equivalent to about 2.04 kg (4.5 lb) of mixed liquor COD (0.68 kg ML COD/kg MLVSS, or 1.5 lb ML COD/lb MLVSS). From Table 9, if oxygen is supplied at a rate of 20.4 kg/day (45 lb/day), the 2.04 kg (4.5 lb) of ML COD and equivalent oxygen is equal to 10 percent of the total mass of oxygen supplied. This 10 percent error in the oxygen utilization efficiency calculation is critical when 90 percent efficiencies are expected. It is possible that greater errors than the previous example in the MLVSS inventory can occur, especially in the clarifier VSS inventory calculation. Calculated values for the change in system VSS inventory (or COD_{Accum}) were a major source of error and contributed greatly to the wide range of oxygen usage efficiency shown. In view of this discussion, the third alternative is not recommended for calculating oxygen utilization efficiency.

TABLE 9. OXYGEN UTILIZATION EFFICIENCIES USING COD BALANCE

Test No.	Length of Test (days)	SI COD (lb/day)*	FE COD (lb/day)*	WAS COD (lb/day)*	Inventory COD (lb/day)*	COD Removed (lb/day)*	Oxygen Supplied (lb/day)*	Percent Oxygen Utilized+
1	10	79.4	15.1	35.9	(27.4)∓	55.5	54.9	101.0
2	13	55.0	8.5	25.5	(1.1)	22.1	44.9	49.2
3	13	71.7	9.2	31.2	(9.4)	40.7	46.1	88.3
4	14	46.2	5.9	22.8	11.3	6.2	35.3	17.6
5	12	48.9	5.5	25.6	56.7	(38.9)	33.6	(115.8)

* lb x 0.4536 = kg

+ Percent oxygen utilized = (COD_R/O_2 supplied) x 100

∓ () indicates Negative Number.

The established criterion for evaluating oxygen treatment systems include determining the oxygen utilization efficiency for that system. Open tank systems cannot readily capture and analyze the off-gas. If it is desired to verify a 90 percent oxygen utilization efficiency for an open tank system, the following approach is recommended:

1. Evaluate mass of oxygen supplied per mass of BOD_5 removed with respect to the operating F/M ratio. Compare this BOD_5 specific oxygen requirement with an expected value. If the calculated BOD_5 specific oxygen requirement is less than the expected requirement value, then the oxygen utilization efficiency must be at least 90 percent. If the calculated requirement is greater than the expected value, then the oxygen utilization efficiency may be less than 90 percent. For this case a second test is recommended.

2. Cap or cover all or a portion of the oxygenation basin. Capture and analyze basin off-gas volume and oxygen concentration. Calculate the mass of oxygen lost in the off-gas, and compare to the oxygen supplied to the system to determine oxygen utilization efficiency.

SLUDGE PRODUCTION ANALYSIS

Earlier investigation of closed basin, high-purity oxygen-activated sludge treatment concluded that significantly less sludge was produced using an oxygen-activated sludge system as opposed to a parallel operated air-activated sludge system.[2] Ball & Humenick[1] took issue with this conclusion pointing out that a truly comparative analysis was not performed because of basic differences in air versus oxygen-activated sludge operations. One difference was the mean sludge age, 2.4 days for the air system compared to 5.4 days for the oxygen system. Ball & Humenick concluded that there may be no difference in cell yield coefficients between air and oxygen-activated sludge and that no study had been performed in which a rational comparison had been made.

The open tank pilot plant oxygen-activated sludge study did not include a parallel operating air system. This section will discuss the sludge yield coefficients derived from the results of the pilot plant operation. These coefficients will be compared with air activated sludge coefficients obtained from the literature. The basic kinetic model is as described by Jenkins and Garrison,[8] Pearson,[9] Lawrence and McCarty,[10] and McKinney.[11] No attempt is made to formally present the theoretical basis for the kinetic model. The terminology for the parameters in the kinetic equation was taken from Jenkins and Garrison.

Basic Kinetic Equation

$$\frac{1}{\theta_c} = Yq - K_d$$

where:

Y = $\frac{\text{mass cells formed}}{\text{mass substrate removed}}$ = yield coefficient

q = $\frac{\text{mass substrate removed/day}}{\text{mass cells in system}}$ = substrate removal rate

K_d = $\frac{\text{mass cells lost/day}}{\text{mass cells in system}}$ = decay coefficient

$\frac{1}{\theta_c}$ = $\frac{\text{mass excess cells/day}}{\text{mass cells in system}}$ = net growth rate

Many investigators have used the basic kinetic equation in a variety of ways; some using substrate as BOD_5, others using substrate as COD. Cell inventory has been used as VSS in the aerator, and as VSS in the aerator plus VSS in the clarifier. In most instances, substrate removed is calculated using the total BOD_5 or COD applied and soluble BOD_5 or COD discharged. Since most literature comparisons use effluent soluble BOD_5 and COD and these measurements were not made, assumptions of values were calculated based on the following relationship. The soluble portion of effluent total BOD_5 or COD was assumed to vary with the F/M ratio. At the highest operating F/M ratio of 1.4 kg BOD_5/day/kg MLVSS, the soluble portion was assumed to equal 90 percent. Similarly, at a lower F/M ratio of 0.7 kg BOD_5/day/kg MLVSS, the soluble portion was assumed to equal 10 percent. The effect of these assumptions on the kinetic equation was to increase the substrate removal rate at a given net growth rate. The assumptions made may be extreme, especially with respect to effluent soluble COD. It should be noted that the greater these extremes, the greater the yield and decay coefficients.

For this report, four alternative methods of presenting the kinetic equation have been graphically illustrated (Figures 12, 13, 14, 15). Figure 12 shows the net growth rate opposed to the BOD_5 substrate removal rate and Figure 13 shows the system net growth rate compared to the COD substrate removal rate. Both include the total mass of VSS in the system. Figures 14 and 15 plot the same parameters as Figures 12 and 13, except the net growth rates included only the mass of VSS in the aerator. The mass of excess cells per day included VSS lost in the clarified effluent and VSS intentionally wasted.

The straight-line relationship in the kinetic equation for sludge production yield and decay coefficients were calculated for the pilot plant and then compared with the coefficients from referenced literature. Table 10 shows one of these comparisons. In one instance, the yield and decay coefficients of the air activated system were less than those of the oxygen

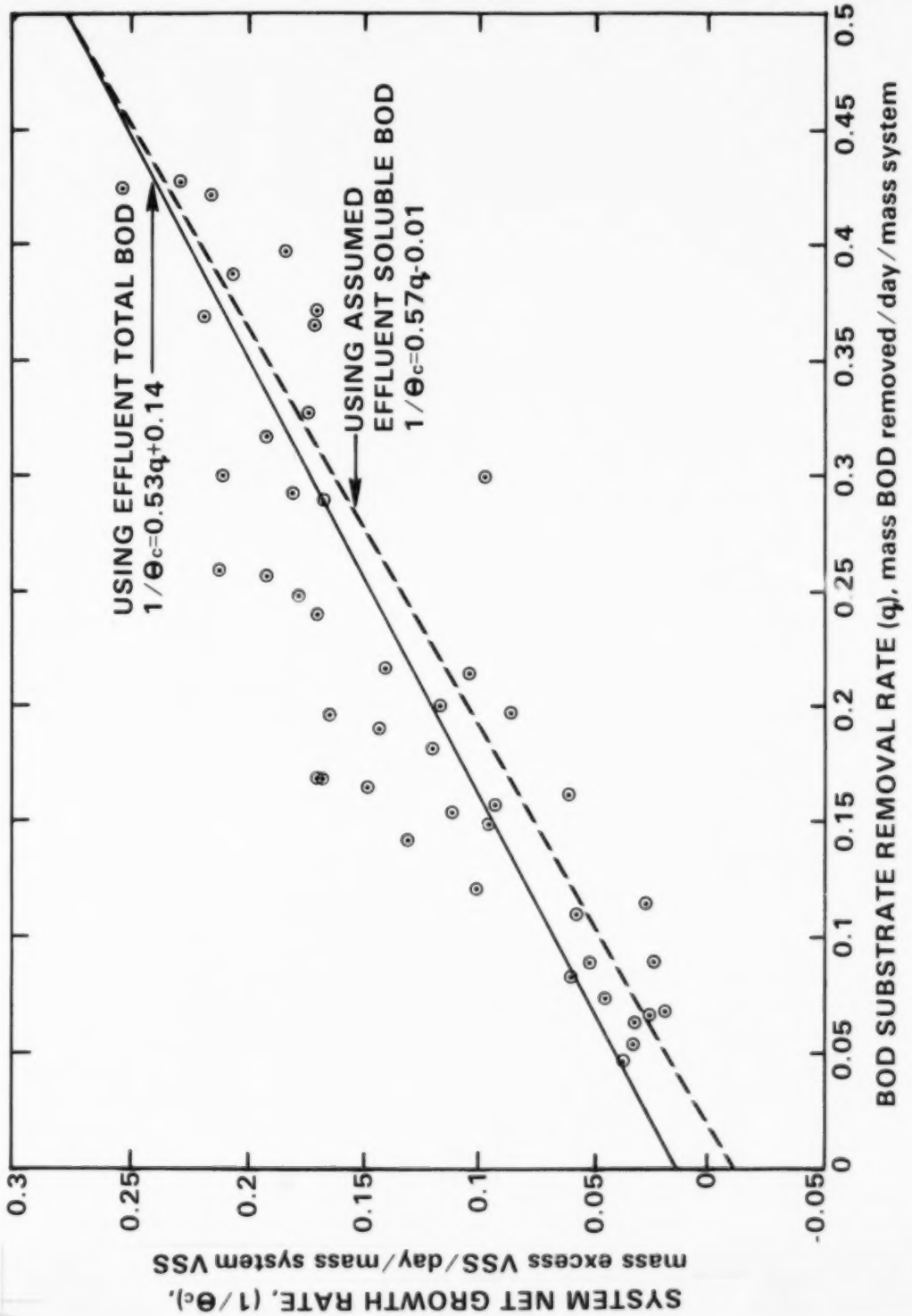

Figure 12. System net growth rate versus BOD substrate removal rate.

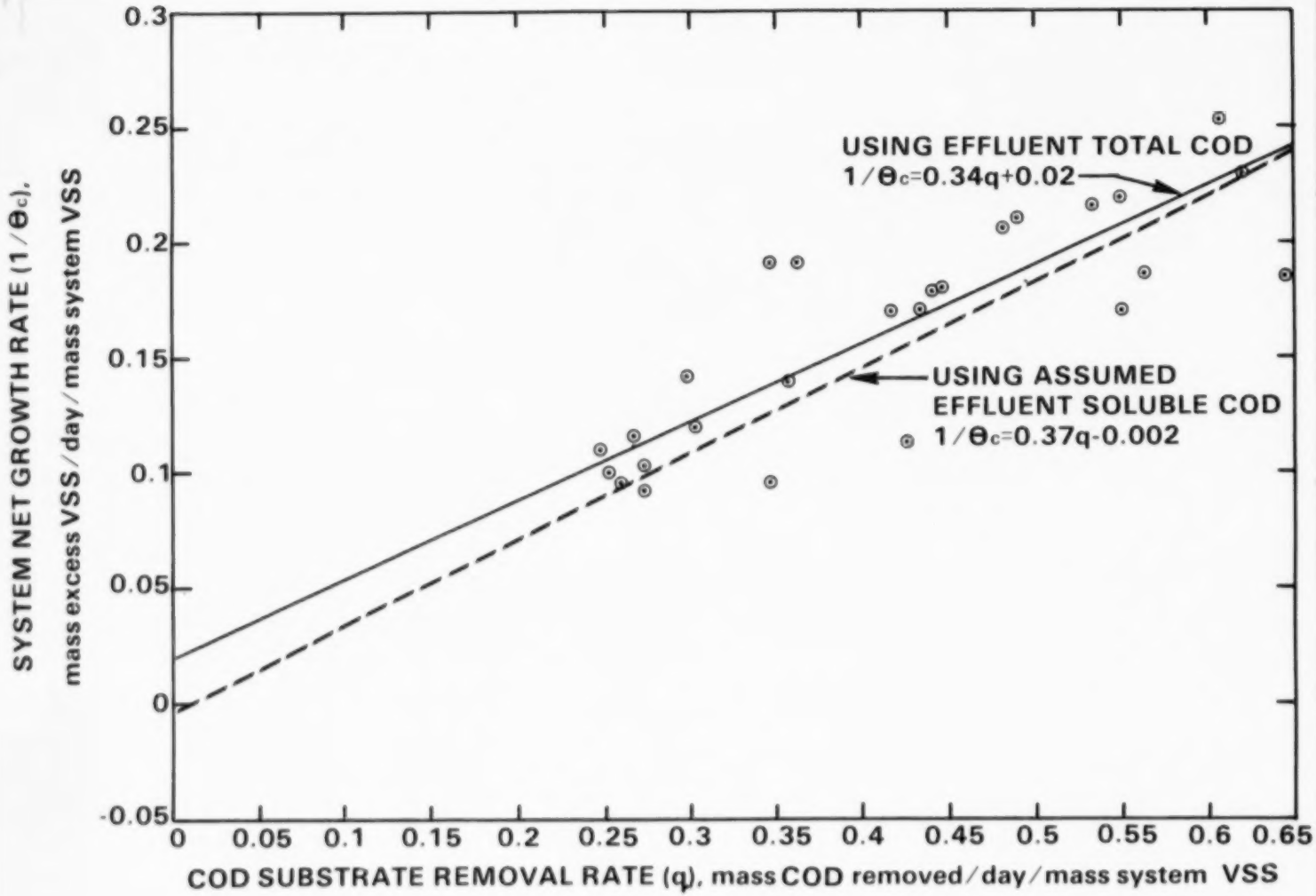

Figure 13. System net growth rate versus COD substrate removal rate.

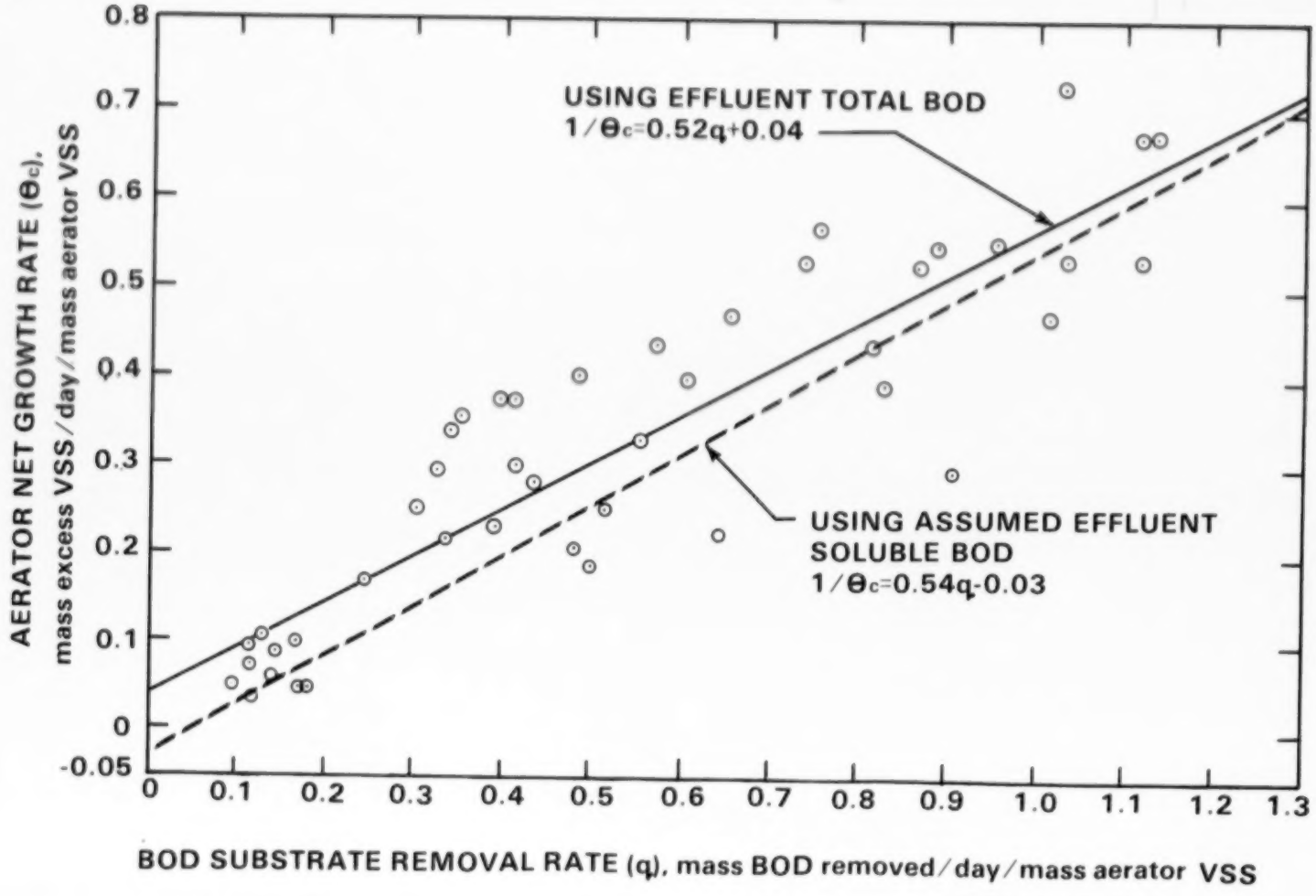

Figure 14. Aerator net growth rate versus BOD substrate removal rate.

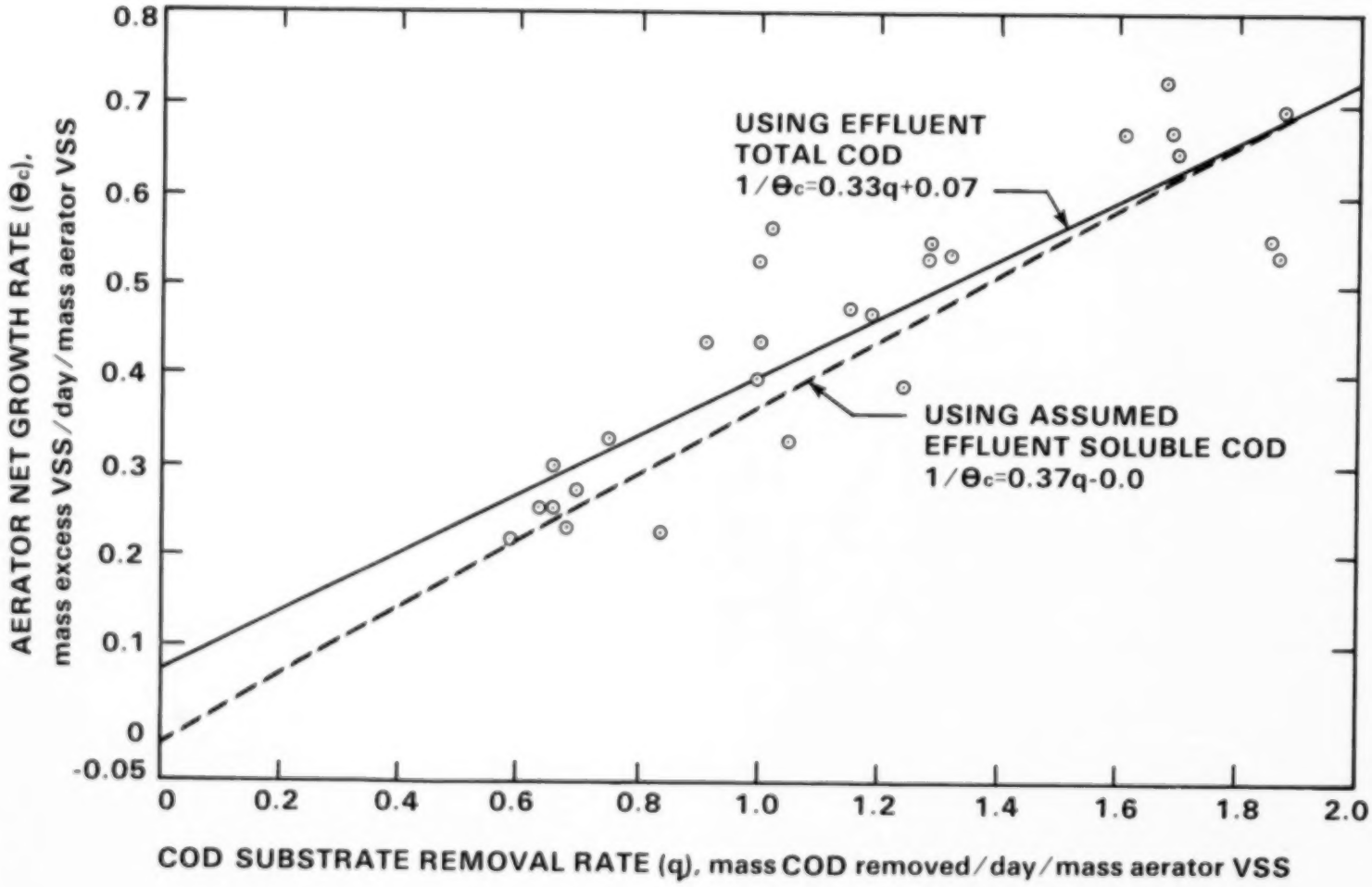

Figure 15. Aerator net growth rate versus COD substrate removal rate.

activated system. In another instance, this situation was reversed. Therefore, further analysis must be made in order to make a definite statement relative to the significant differences in sludge production with air or oxygen activated sludge systems, Yield and decay coefficients obtained were compared with two sets of air coefficients: the Hyperion coefficients reported by Smith and Eilers[12] and referenced by Brenner[13], and typical coefficients reported by Eckenfelder and Ford.[14] The Hyperion plant is an activated sludge plant with a consistent record of excellent performance and adequate mixed liquor D.O. Figure 16 illustrates the relationship between air and MAROX oxygen yield and decay coefffients. It was found that the air yield coefficients were greater and the air decay coefficients were less than the MAROX coefficients. The combined effect of the differences in the yield and decay coefficients on a total sludge production value is defined by the intersection of the air curves with the oxygen curve. The Hyperion example shows a substrate removal rate of 0.2 kg BOD_5 removed/day/kg MLVSS (0.2 lb BOD_5 removed/day/lb MLVSS). The Eckenfelder and Ford example shows a substrate removal rate of 0.51 kg BOD_5 removed/day/kg MLVSS (0.51 lb BOD_5 removed/day/lb MLVSS). In both examples, it was shown that less sludge is produced using an oxygen system at higher substrate removal rates or higher operating F/M ratios. The difference in sludge production is significant when comparing the Hyperion air with the oxygen system. No significant difference exists when comparing the results of the Eckenfelder and Ford air system with the oxygen system.

TABLE 10: COMPARISON OF AIR AND OXYGEN SLUDGE YIELD AND DECAY COEFFICIENTS*

Source+	System	Reference	Yield Coefficient (BOD_5)	Yield Coefficient (COD)	Decay Coefficient (BOD_5)	Decay Coefficient (COD)
MAROX Pilot	Oxygen	Figure 12	0.57		-0.01	
MAROX Pilot	Oxygen	Figure 13		0.47		-0.002
Jenkins-Garrison	Air			0.33		-0.04
Hegg-Burgeson	Air		0.72		0	

* Domestic waste with primary clarification.

\+ Net growth rate and substrate removal rate calculated using mass of VSS in aerator plus mass of VSS in clarifier.

Figure 17 illustrates the relationship (using COD substrate removal rate) between the yield and decay coefficients of Hyperion air and MAROX oxygen. The same interrelationship or trade-off occurred as noted when the BOD_5 substrate removal rate was used. In this case the Hyperion air system would again produce more sludge at a higher F/M ratio. F/M ratios would have to be higher than those corresponding to a COD substrate removal rate greater than 0.48 kg COD removed/day/kg MLVSS.

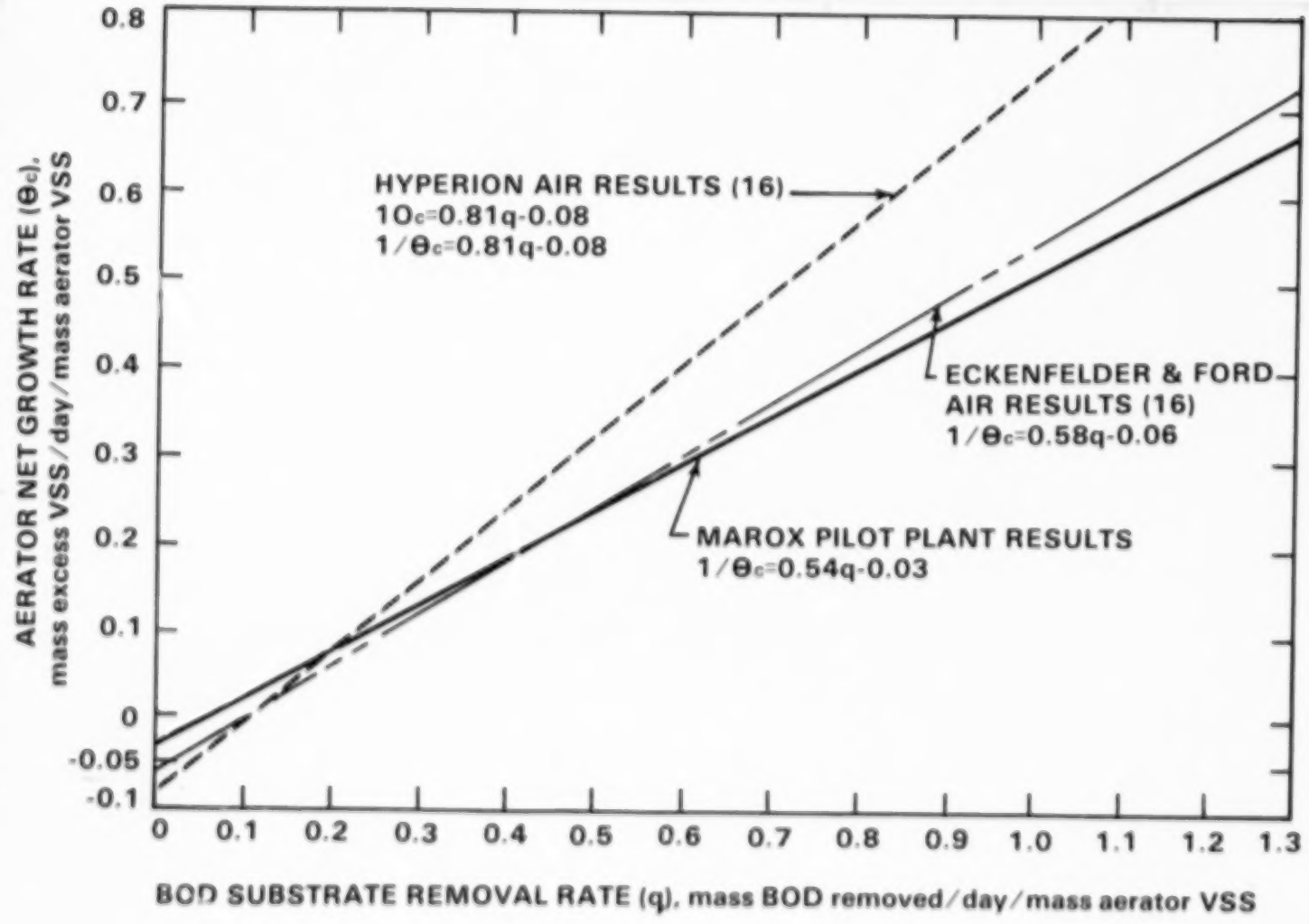

Figure 16. Aerator net growth rate versus BOD substrate removal rate.

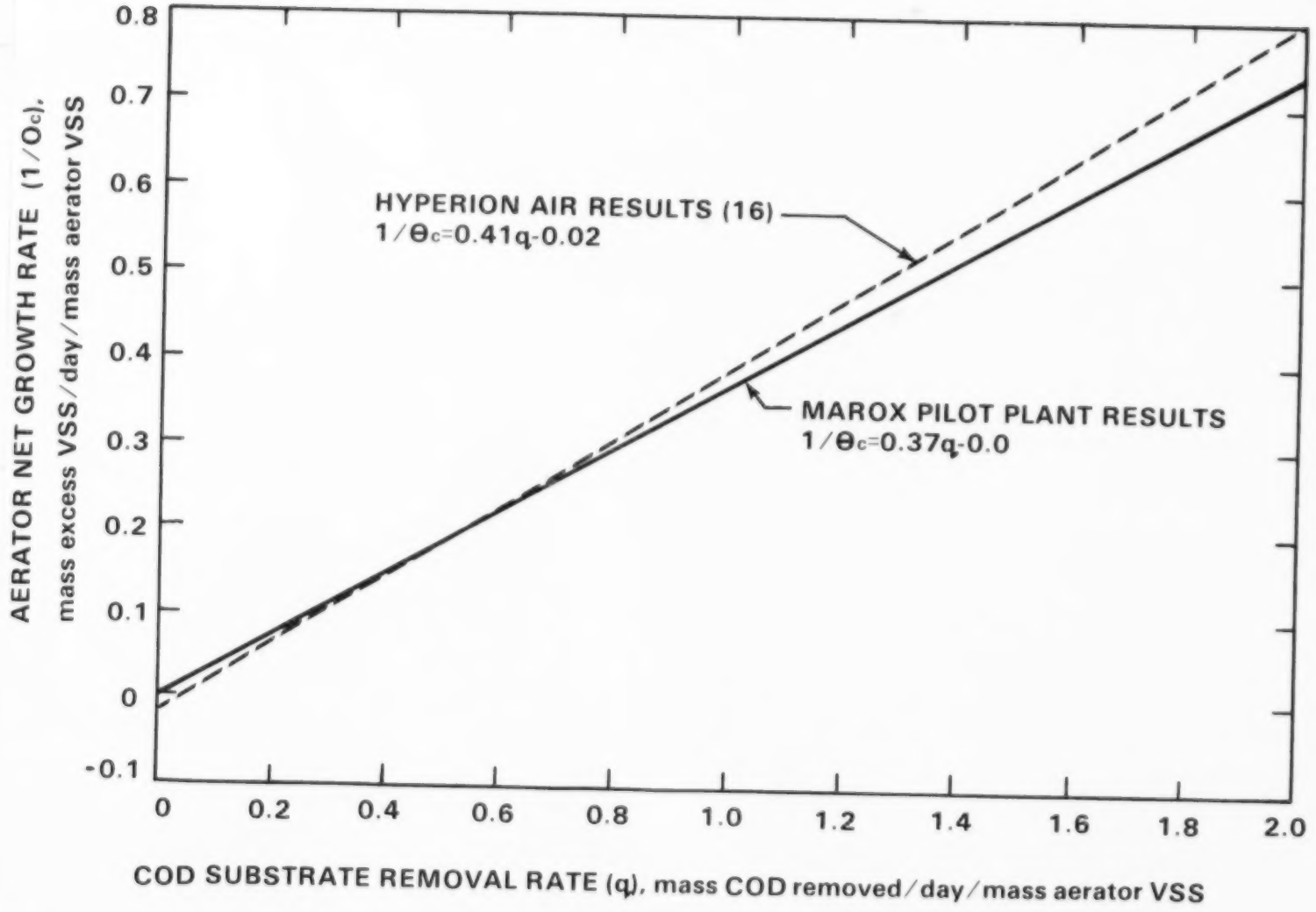

Figure 17. Aerator net growth rate versus COD substrate removal rate.

Based on these analyses and comparisons, it appears that an oxygen activated sludge system at higher F/M ratios will produce somewhat less sludge than the literature-cited air activated sludge systems. One further advantage with an oxygen system is that a higher clarifier underflow concentration is normally maintained with smaller volumes of sludge wasting to maintain a given mass of waste activated sludge. This smaller volume of wasted sludge would reduce the sludge processing requirements for disposal.

SLUDGE SETTLING ANALYSIS

Previous investigation of the use of a high-purity-oxygen closed-tank reactor activated sludge system concluded that an oxygen activated sludge demonstrates better settling characteristics than air activated sludge.[15] The supporting basis for this conclusion was that oxygen sludges demonstrated higher initial settling velocities than air sludges at given TSS concentrations. Initial settling velocities for this test program were determined using a 14 cm (5 1/2 inch) diameter Plexiglas cylinder 8 ft in length. Two longitidunal blades were used to stir the ML at a rate of 10 revolutions per hour to prevent bridging effects of small-diameter cylinders when compared to the large diameters of clarifiers. Figure 18 shows a typical band for air-sludge initial settling velocities at various TSS concentrations, for four different periods of MAROX pilot plant operation. The oxygen sludge in all cases demonstrated good settling characteristics at specific TSS concentrations. A second parameter used to evaluate sludge settling characteristics is the sludge volume index (SVI). During all but two test conditions, the SVI was excellent, averaging less than 85 ml/gm. In the other two test conditions, SVI values were higher than normally expected, but the MLSS initial settling velocities were 1.22 m/hr (4.0 ft/hr) and 1.53 m/hr (5.0 ft/hr). These settling velocities are much higher than typical air sludge velocities for similar TSS concentrations and for those cited in lieterature. At the same time, excellent final effluent quality was achieved even at clarifier solids loading rates as high as 220 kg TSS/day/m^2 (45 lb TSS/day/ft^2 and clarifier overflow rates as high as 33.49 m^3/day/m^2 (822 gal/day/ft^2).

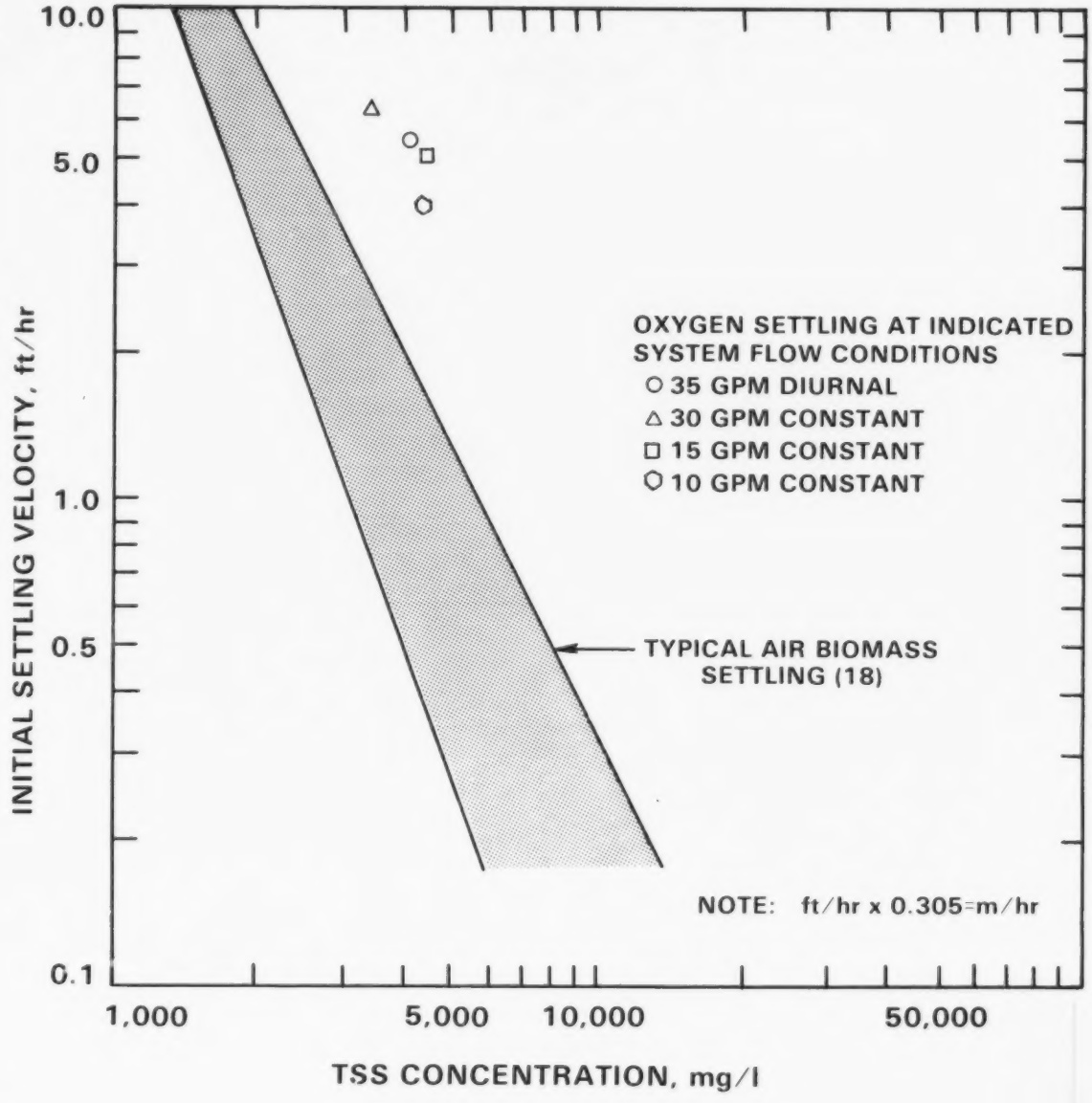

Figure 18. Initial settling velocity versus TSS concentration.

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APPENDIX A

OXYGENATION SYSTEM FOR ACCELERATED SEWAGE TREATMENT (1)

The following is extracted from the literature cited above:

OXYGEN TRANSFER

Oxygen may be transferred to water by the process of diffusion (2). The driving force of diffusion is the concentration gradient between the saturation value and the actual concentration of oxygen in water. However, the saturation value of oxygen in water is related to the partial pressure of oxygen in the atmosphere in contact with the water at a given temperature. For a slightly soluble gas, which does not react chemically with the solvent, the solubility of the gas in a liquid at a given temperature is directly proportional to the partial pressure of that gas. Therefore, if oxygen-enriched air or pure oxygen is provided for aeration, the driving force for oxygen transfer will be increased with the increased oxygen partial pressure, and the oxygen transfer rate will be raised.

The rate of oxygen transfer, dC/dt, is a function of the deficit (D) between the oxygen saturation value (C_s) and the oxygen concentration at any given time (C_t). The relationship between dC/dt and D can be expressed in a general equation:

$$dC/dt = K_La(C_s - C_t) = K_LaD$$

where:

C	=	concentration of oxygen in water
dC/dt	=	rate of change of this concentration
K_La	=	overall oxygen transfer coefficient.

The parameter, K_La, is a function of the interfacial area, volume of liquid, and other physical and chemical variables of the system. The effect of temperature on the value of K_La is ususally expressed as

$$K_La_{(t)} = K_La_{(20^oC)} (\theta)^{(T-20)}$$

where:

T	=	temperature, deg C
θ	=	temperature coefficient which varies from 1.016 to 1.047

$K_L a(t)$ = transfer coefficient at temperature T

$K_L a(20^oC)$ = transfer coefficient at 20^o C

However, it has been reported by Liu et al. (1), that the effect of temperature on $K_L a$ is not apparent over the temperature range 18 to 40^o C; they also concluded that pH does not have a significant effect on oxygen transfer from air to water.

The effects of a number of factors on the rate of oxygen absorption in water have been previously investigated by many researchers. While the facilities used for these studies of oxygen transfer have varied both in shape and size, most of the investigators reached common conclusions on the effect of factors, such as gas bubble size, depth of submergence, contaminants in liquid, gas flow rate, and agitation of liquid.

The general findings are these if all other factors remain constant:

1. The reduction of bubble size will increase the oxygen absorption rate.

2. The oxygen transfer coefficient for a given system decreases with increasing liquid depth; consequently, the rate of absorption is a maximum at the point of bubble formation.

3. The presence of contaminants may increase or decrease oxygen absorption depending on the manner in which the contaminants effect the ability of the gas molecules to penetrate the interfacial survace.

4. The greatest oxygen transfer efficiency for a given system can be achieved at a gas flow rate within a certain range to be determined experimentally. Higher rates do not increase the total oxygenation capacity.

5. The rate of oxygen transfer increases rapidly with the increasing rate of stirring or mixing.

DIFFUSER TECHNOLOGY

Important variables involved in establishing "standard" conditions for testing diffuser efficiencies include (3):

1. Absorbing liquid
2. Method of measuring, computing and reporting rate of oxygen uptake
3. Temperature of absorbing liquid
4. Geometry of test tank and depth of submergence

Figure 1 is a graph depicting oxygen bubble diameter versus liquid depth in clear water which was developed by calculation and checked empirically at MMA. Any combination of bubble diameter and water depth falling below the diagonal line results in the oxygen bubble being completely absorbed, while above the line a part of the oxygen bubble escapes as free oxygen. Thus, it can be seen that for a bubble to be absorbed in 10 ft of water, the bubble diameter must be 0.16 mm or less.

With this criteria in mind, two diffusers were developed: (a) a "passive diffuser" that produces bubbles having an average diameter of 0.2 to 0.3 mm with a maximum of 0.5 mm and (b) an "active diffuser" that produces bubbles having an average diameter of 5 to 30 microns with a maximum of 100 microns. A detailed discussion on the program for diffuser development is beyond the scope of this paper as it involved approximately 3 years of testing and around $700,000 in direct expenditures; however, the key factors of diffuser design will be summarized.

In the passive diffuser, the small bubbles are generated by gas effusion from an immersed porous medium. The desired goal for this type of gas diffuser is to produce minimum diameter bubbles of uniform size which are uniformly distributed over the porous surface and release with uniform rates of formation. Some of the key factors which contribute to achieving the desired goal using the porous medium are: the manner in which the bubble intersects the surface or contact angle, the wetability or surface tension for the porous material, the homogeneity of the porous material throughout its bulk, the liquid depth at which the bubble is formed, and the physical and chemical properties of the liquid in which the bubble is formed. A photograph of several racks of "passive diffusers" ready for installation in aeration tanks is shown in Figure 2.

In the active diffuser, the micro-bubbles are generated by shearing the gas bubble from the surface at the instant it is being formed. The shearing force is produced by liquid flow across the point of bubble formation as shown in Figure 3. The result of the liquid shear is to reduce the bubble diameter about one order of magnitude less than the passive diffuser. The active diffuser also increases the rate of gas or oxygen uptake, by creating a mixing action in the liquid as the bubbles are generated. The key factors in achieving bubble quality with the active diffuser are: gas pressure and gas flow rate, liquid velocity at point of bubble formation, geometry of orifice, diameter of pore and relationship to other pores, liquid depth and the chemical and physical properties of the liquid.

Literature referenced in this Appendix is listed below:

1. Liu, M.S., Branion, R.M.R., and Duncan, D.W., "Oxygen Transfer to Water and Sodium Sulfite Solutions," Journal of Water Pollution Control Federation, Vol. 44, 1972, p. 34.

2. Pai, C.M., "Laboratory Studies Using Diffused Oxygen and Air for Hypolimnion Reaeration," A Thesis, University of Missouri, Rolla, pp. 7-10

3. Morgan, P.F. and Bewtra, J.K., "Air Diffuser Efficiencies," Journal of Water Pollution Control Federation, Vol. 32-10, 1960, p. 1048.

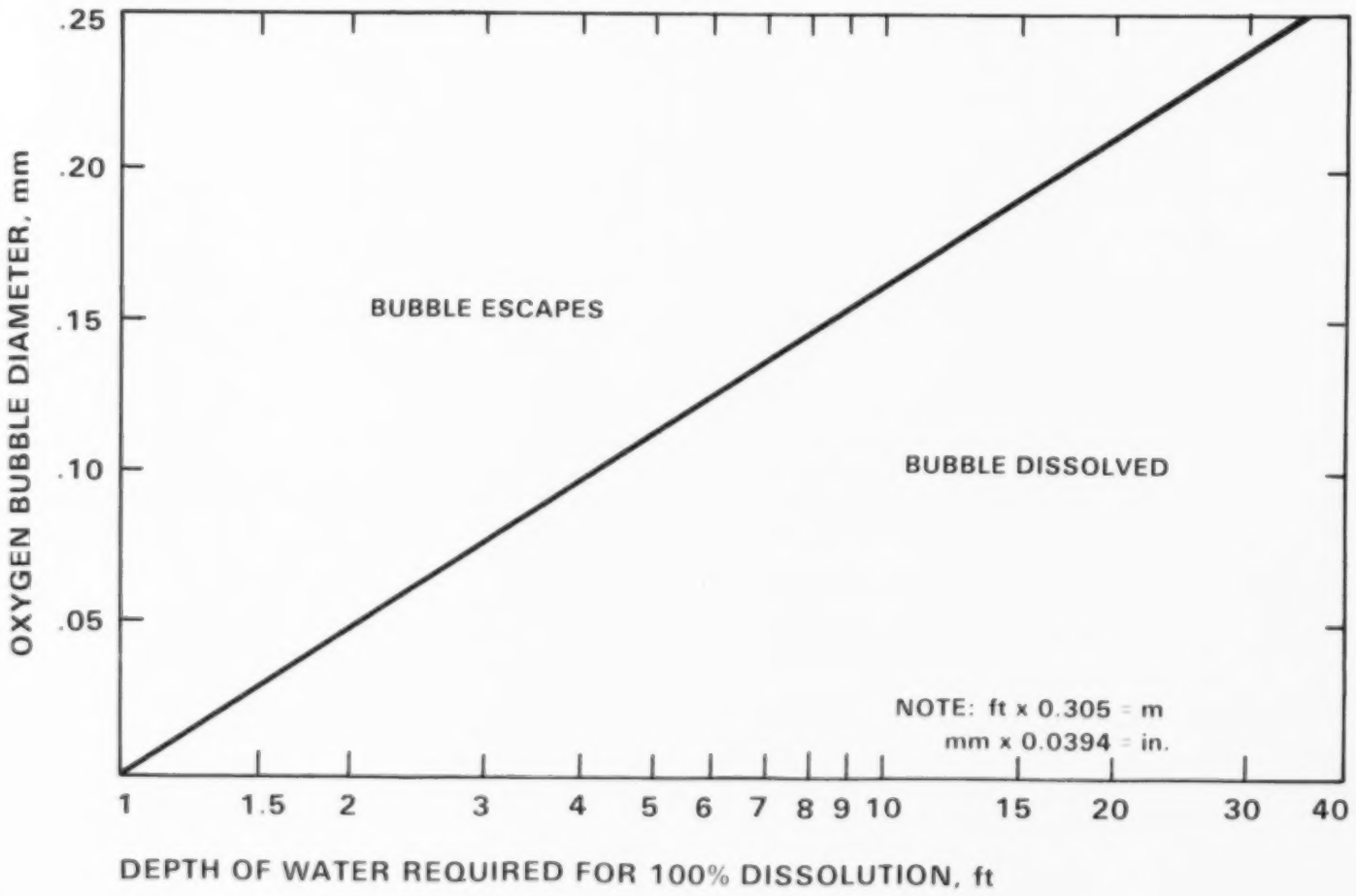

Figure A-1. Oxygen bubble diameter versus water depth.

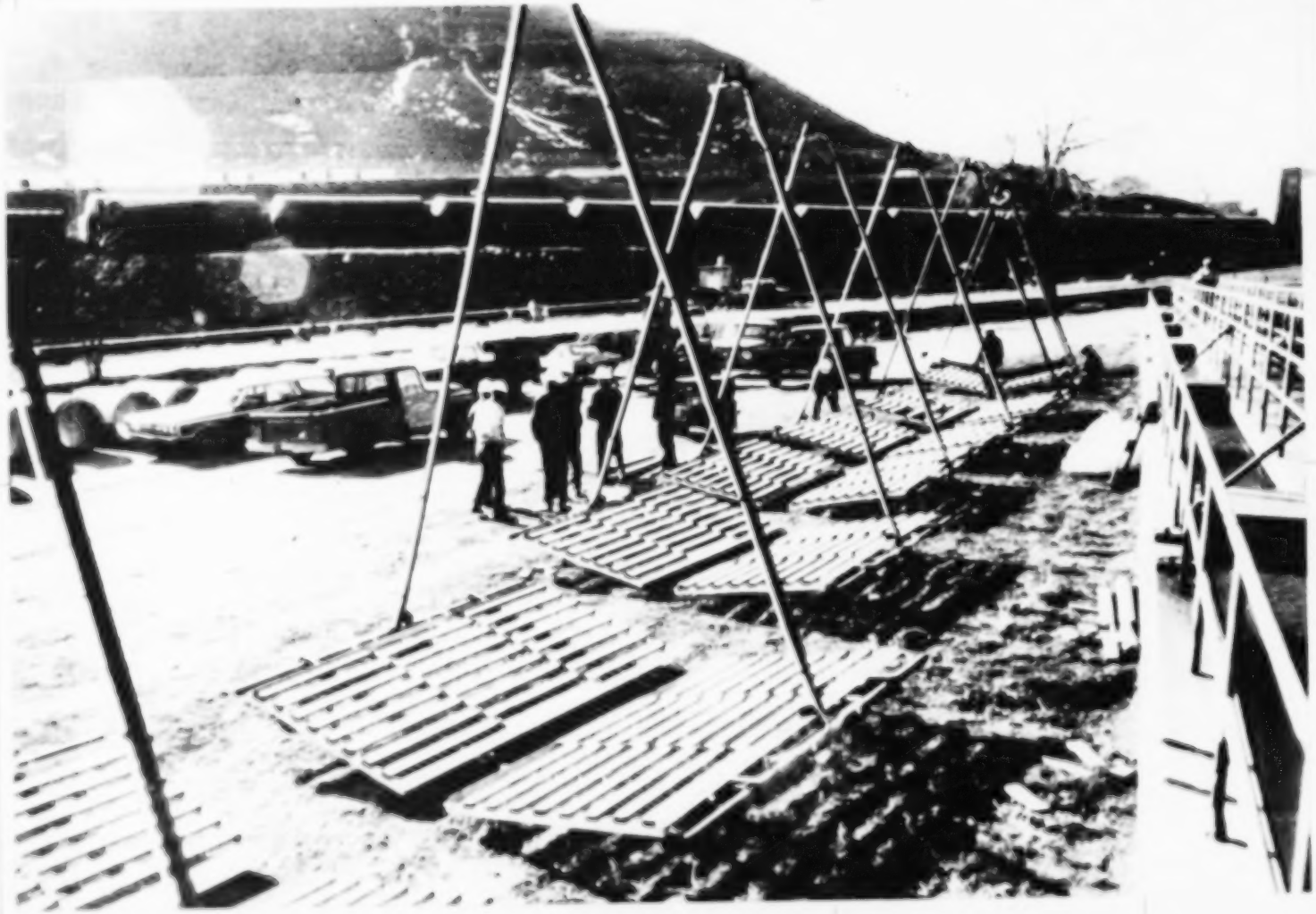

Figure A-2. Passive diffuser ready for installation.

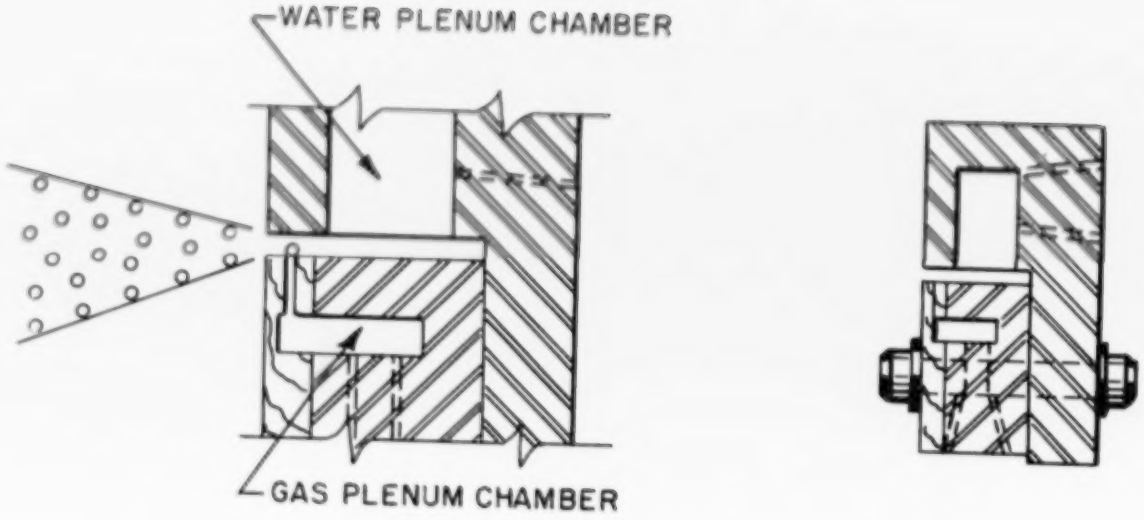

Figure A-3. Active diffuser.